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ASSESSMENT OF THE NEW ZEALAND FOREST RESEARCH INSTITUTE (FRI) PROCESS FOR PINE WOOD HYDROLYSIS AND FERMENTATION TO ETHANOL

FOR

ENERGY, MINES AND RESOURCES CANADA

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c 1983

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TABLE OF CONTENTS

1.0	SUMMARY AND CONCLUSIONS	1
	1.1 Summary	1
	1.2 Conclusions	2
2.0	OBJECTIVES AND BACKGROUND	7
	2.1 Objectives	7
	2.2 Background	9
3.0	DESCRIPTION OF THE FRI PILOT PLANT	14
	3.1 General Process Description	14
	3.2 Process Flow Diagram:	20
	Hydrolysis Section	20
	Preparation for Fermentation Section	22
	Fermentation - Distillation Section	22
	3.3 Operating Procedure:	23
	General	23
	Hydrolysis	24
	Preparation for Fermentation	27
	Fermentation and Distillation	28
	3.4 Results	30
	3.4.1 Radiata Pine:	30
	Alcohol Yields	30
	Sugar Concentration	32
	Total Mass Flow	32
	Energy Balance	35
	3.4.2 Poplar:	36
	Alcohol Yields	36
	Lignin	38
	Energy Balance	38

3.5 Discussion of the Technology 38

3.5.1 Strengths: 39

Experience 39
Yields 40
Wood Preparation 40
Overall Design and Potential for Improvement 40

3.5.2 Weaknesses which can be Corrected: 41

Experience 41
Corrosion 41
Neutralization, Sulphiting and Heat Economy 42
Acid Distribution in the Reactor 43
Fermentability 44
Effluent Treatment 44
Poplar Hydrolysis 45

3.5.3 Weaknesses Inherent in the FRI Process: 45

Sugar Degradation and Humic Substances 45
Scaling up to Full Plant Design 46
Lignin Quality 47
Batch versus Continuous 48

3.5.4 A Second Generation Pilot Plant: 48

Main Hydrolysate Recycle 48
In-Line Neutralization and Temperature Smoothing 50

4.0 ECONOMIC ANALYSIS 51

4.1 Investment 51

4.2 Production Costs 54

4.3 Profitability and Sensitivity 56

Sensitivity 56
Effect of Plant Capacity 59
Expected Income 60
Other Factors Affecting Profitability 62

5.0 POTENTIAL FOR COMMERCIAL DEVELOPMENT 64

5.1 Requirements for Economic Acceptability 64

Favourable Taxation Policy 64
Sale of Lignin 65
Pentose Fermentation 66
Increase in Pump Price of Gasoline 66
Decreased Feedstock and Operating Costs 67
Alcohol as an Octane Enhancer 67
Reduced Sugar Degradation in Process 68

Ranking of Requirements for Economic Acceptability 69

5.2 Comparison with Other Processes 69

5.3 Poplar Processing Economics 73

5.4 Availability of Cheap Feedstocks 75

6.0 APPENDICES 77

A. Process Flow Diagrams

B. Literature

1.0 SUMMARY AND CONCLUSIONS

1.1 Summary

This assessment includes a description of the present FRI pilot plant equipment and operating procedures, and the results obtained in pilot plant operation with radiata pine and with poplar, which are significantly different. Yield of 274 litres of 95% alcohol, equivalent to 258 litres of anhydrous alcohol, per tonne of radiata pine, is higher than any reported by others, but the concentration of fermentable sugars is only 2%. Alcohol yield with poplar is only 190 to 200 litres per tonne. Fermentation of pentoses would increase alcohol yields from poplar quite significantly and bring them close to the yields now obtained from pine.

This report also contains an extended discussion of the FRI technology, in terms of its strengths, weaknesses which can be corrected, and those which are inherent in the system. It also discusses FRI plans for a significant rebuild of the FRI pilot plant to constitute a second generation pilot plant.

Economic analysis is based upon a modest "full scale" plant of 9.9 million litres anhydrous ethanol per year, as it would cost in Canada now. The analysis includes investment, production costs, profitability and sensitivity. Investment would be Cdn. $10,650,000. Requirements to meet the present pump price of gasoline, without reducing taxes, or profits to producers and distributors, are spelled out and ranked. Potential for commercial development is further discussed by comparison with other processes. In addition, attention

is drawn to recent surveys of availability of cheap feedstocks in Canada which prove to be quite encouraging.

1.2 Conclusions

1. The FRI process for conversion of wood chips and sawdust to ethanol is a derivative of the German Scholler process, which has the longest and most varied industrial experience of any ethanol-from-wood process. The FRI pilot plant has achieved somewhat better yields and efficiencies than have been reported previously for this type of process. Yields of 260-275 litres of 95% alcohol have been reached with radiata pine. This would be equivalent to 245-260 litres of anhydrous ethanol per tonne of pine on a dry wood weight basis. This is to be compared with about 200 to 230 litres per tonne reported by others. Other Scholler-derived processes have no advantages, and perhaps disadvantages, compared with a fully matured FRI process.

2. The fermentable sugar solutions produced by the FRI process are very dilute, 2 to 2.5%. A second generation pilot plant has been designed in which main hydrolysate will be recycled for prehydrolysis. This is expected to raise sugar concentrations to 4 to 5%. This new arrangement is expected to show better heat economy and simplification of process flow.

3. For the base FRI plant, 30,000 litres anhydrous ethanol per day or 9.9 million litres per year, the

investment required is $10,650,000, or $1.08 per annual litre. This can be lowered to about $9 million, or 91 cents per annual litre, by purchasing steam instead of burning lignin. This might be the case alongside an existing pulp mill.

4. Production cost in the base case, with wood at $45 per dry tonne and without return on investment, is 44.9 cents per litre which, at 20% return, becomes 66.2 cents per litre. With $15 per dry tonne wood feedstock, this is lowered by 12 cents per litre to 54.2 cents per litre.

5. Larger plants enjoy lower capital and operating costs. A plant to make 120,000 litres per day, a fairly common size now, has an investment of 64 cents per annual litre (down from $1.08 in the base case) and a still larger plant, to make 240,000 litres per day, has an investment of 50 cents per annual litre. This larger plant, with wood at $15 per tonne, has a production cost of 27.9 cents per litre without return on investment, or 37.9 cents per litre with 20% return on investment.

6. Federal and provincial taxation policies can affect viability for these plants. In the base case, lowering combined federal and Ontario provincial gasoline tax by 4.1 cents per litre, which is 25% of the taxes now collected at the pump, would bring the alcohol price in a 10% gasohol mixture in line with the regular gasoline pump price. Any increase in pump price of gasoline would be helpful both from a taxation and from an alcohol viability point of view.

7. Successful pentose fermentation is a near-term prospect and would lower alcohol production cost by 5 to 6 cents per litre (pine).

8. A large scale market for lignin at 25 cents per lb., or 55 cents per kg could reduce the alcohol selling price by 12 cents per litre. FRI process lignin, as now made, may not be suitable for this market.

9. It has been calculated that the use of alcohol as octane enhancer would effect significant refinery operating cost savings. However, with refineries operating at below capacity or shutting down, the oil companies are not rushing to adopt this technology.

10. Ethanol yields from poplar were substantially lower than those from pine, being about 185 litres per tonne. Processing of poplar gave rise to mechanical problems in the degradation and handling of the poplar lignin which interfered with the operation of the pilot plant. Economic calculations resulted in a 20% capital cost disadvantage for a plant based on poplar compared to pine, that is a plant to make 9.9 million litres of ethanol per year would cost $12,780,000 or $1.29 per annual litre (compared to $10,650,000 or $1.08 per annual litre); and that production cost would be 55.7 cents/litre (compared to 44.9 cents/litre. The availability of a practical process for pentose fermentation would be expected to bring the costs of ethanol from the two types of feedstocks closer together.

11. There have been a number of operating problems which are a serious cause for concern. One of these is the formation of large quantities of degradation products of sugars and lignin, forming tars which deposit in pipes and valves, and humic substances which impart colour and perhaps fermentation inhibition to the sugar solutions. Another major problem is related to corrosion. The Incoloy 825 used for the reactor has shown some corrosion, enough to cast doubt on its suitability for a full scale plant. A new second reactor which will be teflon-lined is being ordered.

12. A treatment of the wood sugars to enable them to be fermented satisfactorily has been found and appears to be functioning well.

13. Effluent from the plant has a high COD and BOD. The most satisfactory effluent treatment recommended by FRI is anaerobic digestion to produce methane. The methane so formed, together with the lignin, make the whole plant nearly energy self-sufficient. The capital cost of such a plant is $1.7 million in the base case, with a 2.9 year pay back from the methane produced.

14. Several weaknesses are inherent in the FRI process. The degradation of the sugars and the production of tars and humic substances is one of these. There is also doubt about the degree to which the process can be scaled up without incurring extra costs due to repeated equipment modules. The quality of the ligning produced by the FRI process is such that it is probably only useful as fuel.

15. No other process which has reached the pilot plant operating stage offers advantages over the FRI process. It is considered probable that the two stage steam explosion-plug flow reactor process will prove to be cheaper and better, but no pilot plant based on such a process is yet in operation. Should cheap enzymes become available, they would have significant advantages over acid hydrolysis processes, but no economically viable enzyme hydrolysis process has yet been demonstrated.

16. It has been reported by responsible authorities that the amount of cheap feedstocks available in Canada are sufficient for over 500 plants of the capacity of the 30,000 litres per day plant discussed in this report. Together they would provide 20% of annual motor fuel consumption. They would cost about $4 billion to construct and would have major social and economic benefits for Canada.

2.0 OBJECTIVES & BACKGROUND

2.1 Objectives

This study was authorized September 7, 1982, and its purpose is to provide a detailed assessment of the New Zealand Forest Research Institute (FRI) process for pine wood hydrolysis and fermentation to ethanol. Based on published literature, information obtained during in-depth discussion with researchers operating the FRI pilot plant, and independent process engineering design evaluation of the FRI process, this report will provide the following information:

- a description and discussion of the FRI process, including process flow diagrams;
- capital investment estimates;
- operating cost estimates;
- profitability estimates; and
- potential for commercial development.

For the past four years, FRI has operated a pilot plant for the conversion of radiata pine wastes to sugars, and their fermentation to ethanol, to be used as a liquid transport fuel. Hydrolysis of the wood carbohydrates is effected by dilute sulphuric acid at elevated temperatures and pressures, the process being derived from the earlier German Scholler process and its offspring, the Madison Forest Products Laboratory process and the Tennessee Valley Authority process.

The FRI pilot plant includes careful attention to engineering detail, materials of construction and operating procedures. Reports of their results have left many with the impression that this is possibly the most advanced process now in operation, and a purpose of this report is to examine that possibility.

There is, as far as we are aware, only one other pilot plant in operation on a similar process, the Inventa in Switzerland, and in Section 5.0, a comparison of the FRI process and the Inventa process, as well as other processes, will be made.

In addition, there are about 40 plants based on the Scholler process in operation in the USSR. About two years ago, the Brazilians decided to base their newly planned wood-ethanol plants on the Russian experience. Their circumstances are different from others; these new Brazilian plants will make both charcoal and alcohol. COALBRA, the Brazilian company building the first wood-based ethanol plant, takes its name from its two products, coke and alcohol. Wood coke, or charcoal, is used in the Brazilian steel industry in the same way as coal coke is used in Canada and all other countries. They need wood coke, so problems which would concern us, such as alcohol yields and wood degradation residues, are not as important to them at this time.

There is an active research program and a monthly journal "Hydrolysis and Wood Chemistry" published in the USSR and available in English translation, which reports advances in this field in that coutry. The world literature on this subject is extensive. The most important references are to be found in Appendix 2,

which lists reports from the N.Z. pilot plant operations, those from U.S. and German experiences, and others. The present report will refer to that literature only where necessary to point to further detail.

2.2 Background

There are many approaches to wood hydrolysis:

- dilute acids, hot
- strong acids, cold
- solvolysis with ethanol, butanol or acetone
- enzymatic hydrolysis
- microbiological processes.

The existence of such a proliferation of processes is a symptom of several underlying difficulties:

- wood contains a mixture of carbohydrates, some of which, the hemicelluloses, are rather easy to hydrolyze, or at least no more difficult to hydrolyze than, say, corn starch, while the cellulose is very difficult to hydrolyze, and the conditions required for cellulose hydrolysis are destructive of the sugar which is the product sought;

- wood is only about 50% cellulose, the remainder being hemicellulose and lignin. The hemicellulose is not really suitable for fermentation to ethanol, although there are many other uses for it including other power solvents

or methane gas. The lignin is suitable only as process fuel.

- the economics of production of ethanol from wood are extremely site-specific, being highly dependent upon raw material and energy costs, and are in the best cases marginal. The point about all the research has been to improve the technology in order to improve the economics, to bring more locations into the profitable range.

Some factors peculiar to the New Zealand situation can be listed:

- an island with few resources, highly dependent upon importation of liquid transport fuels,

- excellent plantation forests of radiata pine with a highly developed forest products industry throwing off substantial quantities of waste. Continued plantation activity has resulted in the expectation that in about 10 years time there will be a substantial surplus of pine timber.

- a highly developed forest research infrastructure including the magnificent Forest Research Institute, and a highly educated and science oriented population.

Offsetting these motivations, New Zealand has a supply of natural gas and has now undertaken to convert this on a very large scale to gasoline using the Mobil process; and the forest products wastes command a good price when exported to other paper-making countries, especially

Japan. The result of all these forces is that the Government of New Zealand sees its best opportunity to sell the FRI process technology abroad, especially to U.S. investors, and negotiations for a U.S. plant to be built in Arkansas are in an advanced stage. Meanwhile, a private consortium of sawmill owners is building a plant to make five million litres of anhydrous fuel ethanol per year at Rainbow Falls, New Zealand, based on the FRI process with some modifications, only a few kilometres from the four-year old pilot plant and scaled up by a factor of 300. This plant will be completed well before the U.S. plant - perhaps early enough that design innovations can be borrowed from it. It is expected to be in operation in January 1983.

Interest in the FRI process has developed rather slowly in Canada. There are several reasons for this:

- We have large stands of poplar which are underutilized and which may form the best raw material for Canadian ethanol-from-wood plants, whereas the FRI process was developed with pine. The behaviour of the two species is somewhat different and the products are somewhat different. We have, of course, large quantities of pine and other coniferous forest wastes. These are either utilized for pulp making, or are increasingly being used for energy by direct combustion, either hogged or pelleted (The Bio Shell process). These processes do not produce liquid transport fuels.

- Canada has not felt any great shortage of liquid transport fuels. Alternative sources of liquid

transport fuels for the longer term have been seen in the tar sands, the heavy oils, the Arctic oils and in natural gas and its products. Thus there has been no sense of urgency in the development of wood-based processes for liquid transport fuels.

- A competing process for wood utilization for liquid transport fuels has been promoted by both federal and provincial agencies in an attempt to develop a methanol-from-wood process. This is difficult to understand in the light of Canada's huge methanol-from-natural gas production, but is a fact of life which has affected interest in ethanol-from-wood.

- Canada's forest industries have been quite cool and even hostile to the whole concept, since it has been seen as a potential competitor for wood supply, thereby threatening to raise the cost of wood. The large oil companies have also been cool, especially since present refinery capacity is underutilized. This is so in spite of Canadian studies which show that ethanol as an octane enhancer confers important economic benefits on refinery operation. The forest industries and the oil companies together represent an important fraction of the Canadian economy, and their arguments are listened to with consideration and sympathy by those decision makers who, in the end, determine the direction of technological development.

In consequence, interest in the FRI process has been slow to develop in Canada. Recently, Canertech, which is planning an ethanol-from-poplar pilot plant, has had a study made of the performance of poplar in the FRI pilot plant. The results were rather poor, but Canertech may still give serious consideration to this process for their pilot plant. Several other Canadian companies - Weston, Mohawk Oil, Stake Technology, Iotech, and others - are considering wood-based or cellulose-based alcohol plants. The FRI process is beyond question one of the options which must be examined in preparing for such a development. Hence the importance of the present study.

3.0 DESCRIPTION OF THE FRI PROCESS

3.1 General Process Description

Following two years of laboratory and literature studies, the pilot plant required two years for design and construction, and was first operational in March 1979. The process is conceived as a two-stage process. In the first stage, prehydrolysis, radiata pine chips and sawdust mixture are treated with hot dilute sulphuric acid at 140-150°C in a pressure vessel - the hydrolysis reactor - for the breakdown of the hemicelluloses of the wood. In the second stage, the main hydrolysis, fresh dilute sulphuric acid solution at a higher temperature, 175-190°C is added while simultaneously draining the lower temperature solution from the bottom of the reactor, to break down the cellulose. Hydrolysis of the cellulose requires about two hours or more at this temperature. The drainings contain sugars from the hydrolysis of first the hemicellulose and later the cellulose. When the hydrolysis is judged to be complete by the low concentration of sugars in the solution being drawn from the bottom of the reactor, the reaction is terminated by adding cold water and finally flushing out the lignin residue from the reactor.

The FRI process is a batch process: one batch of chips-sawdust is treated at a time. Once the wood is in the reactor, it is not handled further. Essentially the wood remains in place while the acid, which moves downward through the bed of chips and sawdust, leaches the sugars out and carries them to the base of the reactor, where the sugar (and acid) solution passes

through a screen as it is withdrawn.

When processing pine, chips retain their shape although not their mechanical strength, and the lignocellulosic residue settles down in the reactor. When processing poplar, the chips lose their shape and the lignocellulosic residue is rather muddy and more difficult to handle.

The single hydrolysis reactor is an insulated pressure vessel, a vertical cylinder 1.9 m high and 0.45 m in internal diameter, with a working volume of 220 litres. It is designed to operate at pressures up to 3000 kPa (435 psig). The reactor body is fabricated of Incoloy 825 for corrosion resistance in the presence of the hot sulphuric acid.

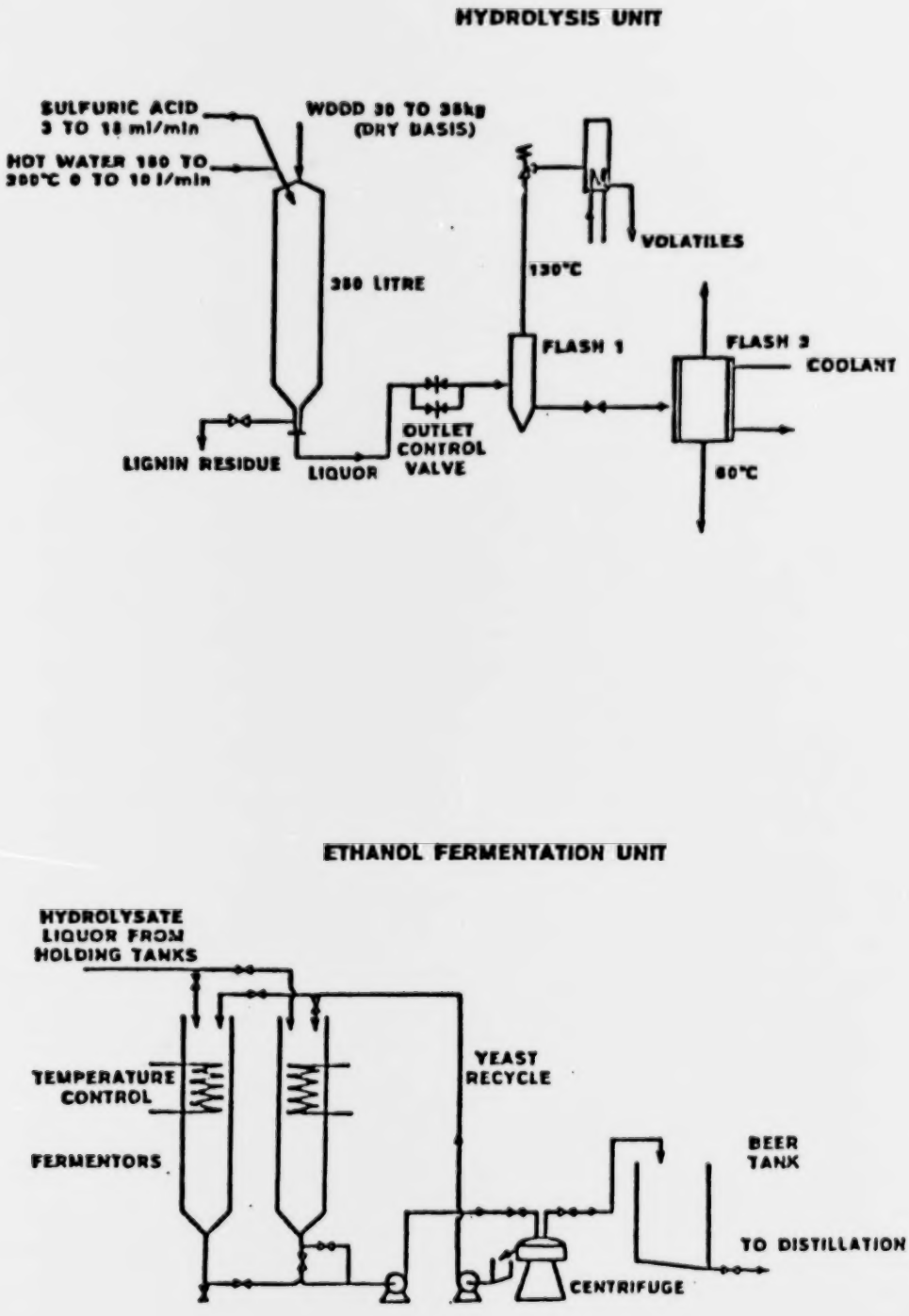
A batch of wood is 70 kg wet weight at 50 to 60% moisture, that is 30 to 35 kg dry weight of wood. A complete cycle from one filling to the next requires about four hours. If operated at capacity around the clock, it could process about 0.2 tonnes per day. For filling with wood, the cover is removed and there are ports for addition of acid at the top, for removal of hydrolysate and for removal of residual lignin at the end of the reaction at the bottom. The liquid strainer is in the form of a vertical cylinder centrally located in the narrowed bottom of the reactor.

Included in the pilot plant is equipment for cooling the hydrolysate while recovering volatiles, for preparation of the hydrolysate for fermentation, and for fermentation and distillation of the alcohol. The schematic diagram in Figure 1, based upon a diagram in

Paper No. 6, Appendix 2, conveys some idea of the general arrangement of the equipment. It shows the reactor as receiving wood and acid at the top, and sugar solution and lignin residue being withdrawn from the bottom. The temperature is shown as high as 200°C, which would generate a steam pressure of 1470 kPa = 213 psig. The sugar solution is removed to a flash tank to lower the pressure and enable flashing off of the volatiles, which are condensed and collected. The sugar solution is stored in tanks which keep separate the prehydrolysate - the early, lower temperature drainings - from the main hydrolysate, the later, higher temperature percolate. There is a treatment stage, indicated but not shown, in which the acid is neutralized with lime. After filtration, the sugar solution is pumped to fermenters to which yeast and nutrients are added and in which alcohol is formed. At the end of fermentation the yeast is separated in a centrifuge for reuse. The dilute alcohol is stored in a beer well, from which it is pumped to the distillation column to separate and concentrate the alcohol. The product is 95% alcohol, which is suitable for use as transport fuel but not for blending with gasoline, which would require a dehydration step.

The pilot plant is highly instrumented, including a computer for data logging, as well as appropriate control instruments. General views of the pilot plant are shown in Figures 2, 3 and 4 and the control panel and some of the instrumentation in Figure 5. Figure 2 shows the hydrolysis reactor, and its general appearance and size. The fermenters are shown in Figure 3, while Figure 4 shows the control panel and the data-logging computer. A larger view of the control panel and some

FIG. 1


HYDROLYSIS UNIT
SULFURIC ACID
3 TO 18 ml/min
HOT WATER 180 TO
200°C 0 TO 10 l/min
WOOD 30 TO 38kg
(DRY BASIS)
VOLATILES
130°C
380 LITRE
FLASH 1
FLASH 2
COOLANT
OUTLET
CONTROL
VALVE
LIGNIN RESIDUE
LIQUOR
60°C
ETHANOL FERMENTATION UNIT
HYDROLYSATE
LIQUOR FROM
HOLDING TANKS
TEMPERATURE
CONTROL
FERMENTORS
YEAST
RECYCLE
BEER
TANK
TO DISTILLATION
CENTRIFUGE

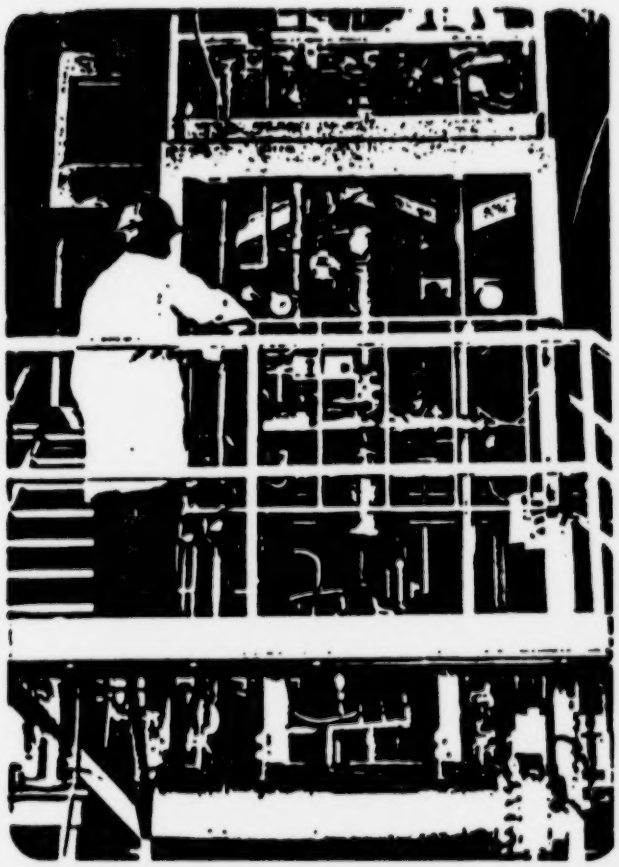


Fig. 2
The wood hydrolysis plant - the main reactor vessel (FRI CS 23739)

Fig. 4
The control room for the ethanol plant (FRI CS 23742)

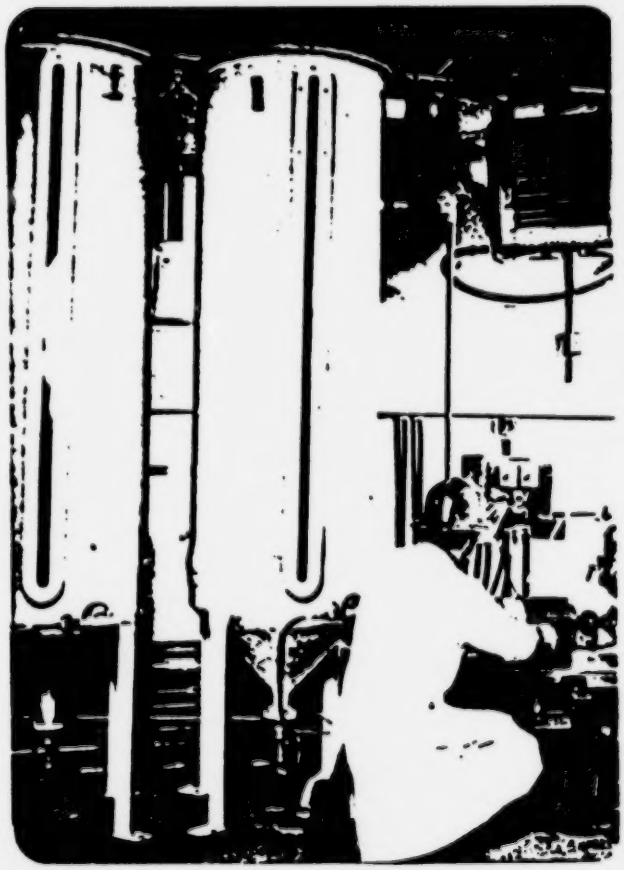


Fig. 3
The wood hydrolysis plant - the fermentation vessels (FRI CS 23743)

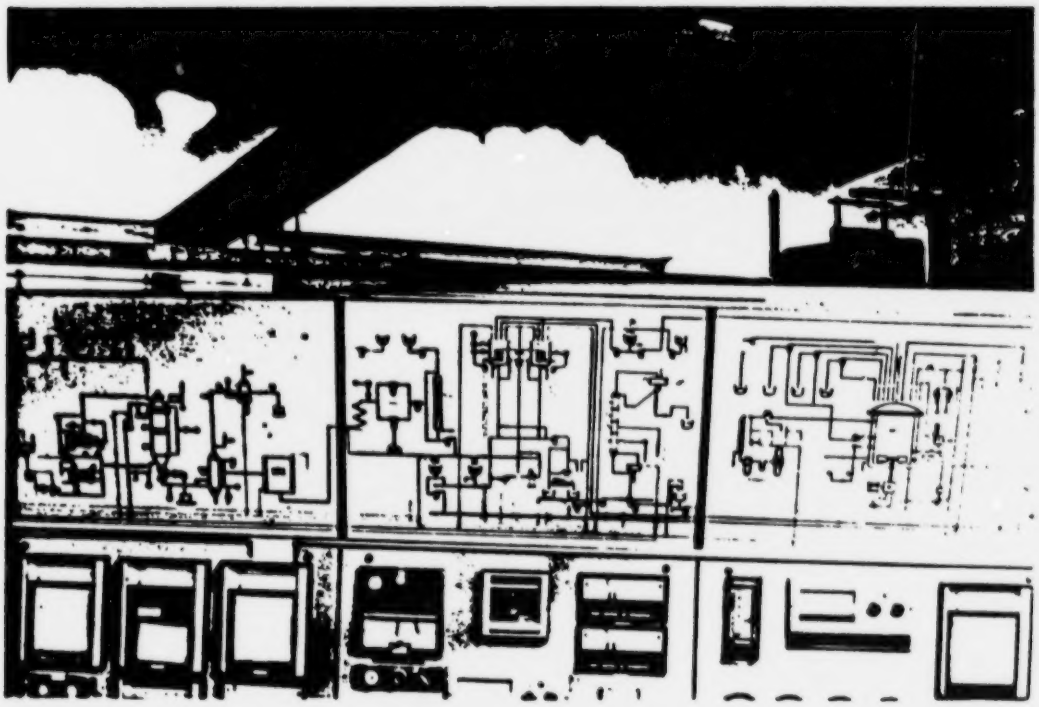


Fig. 5 Control panel schematic flow diagram for the ethanol plant

computer. A larger view of the control panel and some of the instrumentation is shown in Figure 5. This panel has not been updated as the pilot plant has been modified, and the process flow diagram shown in Appendix 1 is a more realistic representation of present equipment and practice.

3.2 Process Flow Diagram

The process flow diagram in Appendix 1 shows all the principal vessels in the pilot plant.

Hydrolysis Section

The hydrolysis reactor is an insulated pressure vessel, pressure rated at 3,000 kPa (435 psig) and normally operated at or below 1,600 kPa (232 psig). The total volume is 250 litres and the working volume is 220 litres. It is a vertical cylinder 1.9 m high, 0.45 m diameter. It is fabricated of Incoloy 825, a high nickel alloy of approximate composition Fe 21%, Cr 22%, Ni 42%, Mo 3%, Cu 3%, Mn 1%, Ti 1%. The cover is removable for charging with wood. The reactor has openings near the top for receiving acid solutions or cold water, and steam inlets near the bottom and top. A strainer is located in the narrow bottom of the reactor, in the form of a vertical cylinder which leads to a 35 mm outlet pipe for hydrolysate. A 100 mm pipe and valve are located at the side of the bottom cone for removing residue.

The dilute sulphuric acid solutions required for hydrolysis are made up by metering concentrated sulphuric acid from its storage tank to a mixing nozzle

located near the top of the reactor, where it is mixed with water preheated to the required temperature. The mixing nozzle is fabricated from solid teflon rod.

Water from a supply tank is first heated with geothermal steam to about 125-140°C - geothermal steam is quite irregular - and then brought to the required temperature by electric heating. The temperatures required may vary from 150 to 155°C for prehydrolysis to 200°C for the main hydrolysis, so both water heating vessels are pressure vessels rated to accept these pressures. They are not designed for acid service, and are fabricated of mild steel. Temperature control is reasonably accurate to $\pm$ 1°C.

Following hydrolysis, the sugar solutions (hydrolysates) are withdrawn from the reactor. First entering a flash vessel where the pressure is reduced to 250 kPa (37 psig), flashing off the volatiles which are condensed by a vapour condenser and collected. The flash vessel is a pressure vessel fabricated of 316 SS. In this operation the temperature falls to 140°C. The hydrolysate is further cooled in an atmospheric pressure flash vessel, the vapours, mainly steam, being allowed to escape. About 12% of the hydrolysate volume is evaporated in the two flash vessels. The final temperature of the hydrolysate is about 90°C.

A residue tank is provided for receiving the lignin, or lignocellulosic residue, remaining in the reactor after hydrolysis.

Preparation for Fermentation Section

Hydrolysate from the cooler is collected in three tanks, all fabricated of mild steel with polypropylene lining. The prehydrolysate tank has a capacity of 250 litres, the main hydrolysate tank is 600 litres and the washings tank is 300 litres. In the hydrolysate tanks, the acid sugar solutions are neutralized, by adding a slurry of either lime or powdered calcium carbonate. In the original design, neutralization was carried out in an in-line mixer shown as a zig-zag line on the control panel shown in Figure 5, the idea being to neutralize without loss of pressure, but this proved impractical and neutralization is carried out at atmospheric pressure. The neutral solutions are then pumped to a mild steel pressure vessel, the sulphiting vessel, in which they are heated to 130°C, and a small amount of sodium sulphite is pumped in. This vessel provides about 30 minutes retention time.

The temperature and pressure of the sulphited sugar solution is reduced in a flash vessel equipped with a cooler to 70-90°C, whereupon it is pumped to a filter, and from there to the fermenters. The filter cake, which is mainly calcium sulphate and other calcium salts plus some tarry humic materials, is collected in a filter residue tank for disposal.

Fermentation - Distillation Section

There are two fiberglass polypropylene-lined fermenters, each 600 litres, equipped with cooling coils for temperature control and agitators. There are small tanks for preparing solutions of nutrients: urea,

diammonium phosphate, and molasses. In addition there are facilities for adding yeast, both fresh dry yeast, and yeast cream from the centrifuge for recycling. An outlet at the top of each fermenter vents CO_2 to the atmosphere.

Upon completion of fermentation, the contents of the fermenters are centrifuged in a Westfalia SAI-02-175 Centrifuge. Yeast cream so obtained is returned for the next fermentation batch. The clarified beer is collected in a beer well from which it is fed to the stripping section of the distilling column. This column, 100 mm in diameter and 5.2 m high, has a lower stripping section and an upper rectification section. It is a packed column, the packing being mini-contact rings of 316 SS. The column has a reflux condenser at the top and a reboiler at the bottom. Stillage, stripped of alcohol, is collected in a stillage receiver, while the 90 to 95% alcohol taken from the reflux condenser is stored in an alcohol receiver. Stillage is used in experiments, the excess being disposed of.

3.3 Operating Procedure

General

Operating procedures have varied over time, as is appropriate in a pilot plant. The following represents operating mode in late 1982. It is not the preferred procedure, but the best possible with the available equipment. Plans for improvements in the near future are discussed in Section 3.4.

The preferred raw materials are chips and sawdust, the mixture having better space filling properties and better acid flow properties than either alone. Hogged waste is also used. Bark is not seen as a problem, say in interfering with fermentation, but merely as non-productive reactor space occupation.

Hydrolysis

The hydrolysis reactor is hand-loaded, with some minor attempt to bat the wood down, but no serious attempt is made to achieve a high digester loading. After loading, the cover is rebolted in place and steam is released into the digester, primarily to vent off air, but also to achieve uniform moisture within the wood charge and to preheat the wood and the digester. Steaming is continued over 15 to 30 minutes until a uniform temperature profile in the reactor at 130 to 135°C is observed. Condensate, amounting to 18 to 20 litres, is not removed. It contains about 0.4 kg of sugars which is about 3% of the total to be expected.

At this time, a carefully prepared mixture of concentrated sulphuric acid and preheated water are added to the hydrolysis reactor. The amount of sulphuric acid is controlled by a metering pump. The temperature of the water, about 150 to 155°C, is controlled by electric heating as already described. The acid and the water are mixed in a teflon nozzle which sparges the acid solution across the bed of wood. Addition of acid is continued until the wood is covered. The temperature at this time increases from about 145°C at the bottom of the reactor to 150°C at the top of the wood charge. The acid concentration is, on the average,

0.45 to 0.5% H_2SO_4 by weight of solution in the reactor.

Filling the reactor with acid requires 20 minutes. During filling, sugars released by hemicellulose hydrolysis are carried from the top of the reactor to the bottom. This results in a "plug" of carbohydrates near the bottom of the reactor, and a high initial concentration when drainage begins. As soon as the reactor is full, drainage of liquor is begun from the bottom of the reactor, and hotter dilute sulphuric acid solution - actually hotter water - is added simultaneously at the top. The mixed acid concentration is 0.5% H_2SO_4. The higher temperature is about 185°C, although it may be 10°C hotter or cooler. The rates of withdrawal of liquor from the bottom of the reactor and addition of acid solution to the top of the reactor are controlled to be the same. In consequence the reactor remains full of liquor and the wood covered with acid throughout the hydrolysis, which is about two hours.

The above procedure involves two important and quite recent departures from standard practice as described in FRI publications even just six months old (3). It had been the practice to allow the 150°C hydrolysis - the prehydrolysis - to proceed for 30 minutes before commencement of drainage of prehydrolysate. Studies of rates of solution of the various components of the prehydrolysate convinced the pilot plant scientists that this waiting period is unnecessary - that the prehydrolysis is very rapid at this temperature. So there is now no waiting before drainage begins. The second change is that until very recently the reactor

was drained of prehydrolysate almost completely before addition of the 185°C acid began. Under those circumstances the reactor contained only wet wood, and the acid solution percolated through the bed. This is now no longer the case: the reactor remains full of liquor until hydrolysis is complete.

The time required for hydrolysis varies with the usual variables, temperature and acid concentration. The rate of acid addition can be varied over a fairly wide range, two to eight litres per minute with normal operation at about four to five litres per minute. Percolation time is about 120 minutes, thereby giving 500 to 600 litres of hydrolysate of about 2 to 2.5% sugar solution. Yields of sugar are somewhat higher with faster flow rates, but then the sugar solutions are more dilute. The outflow of percolate is monitored with a sugar refractometer calibrated with wood sugar solutions. The run was considered complete when the refractometer reading gave zero sugar concentration. Thus all of the sugar is washed from the wood by the hot acid.

At the conclusion of hydrolysis, there remains in the digester the original wood lignin, modified by the reaction. The bed has shrunk in volume and now occupies about one-half to one-third of the reactor. Cold water is admitted to the reactor to wash out the acid, and then the lignin is washed out to a receiver. This is not an altogether smooth operation since the lignin is accompanied by tarry, humic materials which are products of sugar degradation, and reaction products of lignin with the sugar degradation products. The "lignin", however, analyzes as 85 to 95% Klason lignin. It is used for experiments, and the excess disposed of.

Returning now to consideration of the hydrolysate, as hydrolysate is withdrawn from the reactor it enters a flash tank where the pressure and temperature are reduced and volatiles are flashed off. The pressure is reduced from about 900 kPa (130 psig) to 195 kPa (28 psig) and the temperature from 185°C to 140°C in the flash tank. The volatiles, being furfural, methanol and 2-butanone (methyl ethyl ketone, a common solvent), are condensed and collected for analysis and disposal.

The hydrolysate then enters a second flash tank at atmospheric pressure where it is cooled to about 90°C. No attempt is made to collect the steam so generated. In both flash tanks together, about 12% of the volume of the solution is evaporated.

From the second flash tank, the liquor is directed to one of three tanks, the first being for prehydrolysate, the second for main hydrolysate and the third for washings.

Preparation for Fermentation

The hydrolysates are highly acidic, containing sulphuric acid as well as sugars, and in addition several organic acids such as formic acid and levulinic acid formed by sugar degradation. These organic acids account for 30-40% of the total acidity. Lime or limestone are used for neutralizing the acids (CaO, $Ca(OH)_2$ or $CaCO_3$) for reasons of cost, and the resultant precipitates, while mainly $CaSO_4$, gypsum, contain also Ca salts of these organic acids, and humic tars formed from reaction products with lignin. Gypsum is relatively insoluble; however, the presence of the calcium salts of organic

acids increases its solubility in the neutralized hydrolysate.

Neutralization is carried out batchwise in the hydrolysate tanks by adding sufficient lime or powdered limestone to raise the pH to about five. The temperature at time of neutralization is about 90°C. Following neutralization a small amount of sodium sulphite is added, sufficient to give 0.05% Na_2SO_3. The liquor is then pumped into a pressure vessel, the sulphiting tank, in which its temperature is raised to 130°C, and held at that temperature for 20 to 30 minutes. The purpose of the hot sulphiting treatment is to offset fermentation inhibitors in the wood hydrolysates. The beneficial effects of this treatment have been clearly demonstrated in their experience.

The neutralized, sulphited hydrolysates are flash-cooled to near fermentation temperature. They are then filtered, and pumped into the fermenters.

Fermentation and Distillation

Fermentation is carried out with bakers' yeast, Saccharomyces cerevisiae. To the hydrolysates in the fermenter are added small quantities of nutrients, to maintain healthy activity of the yeast. The nutrients normally used are urea and fertilizer grade diammonium phosphate, while some molasses is also added as a source of vitamins and trace elements. Both fresh yeast and recycled yeast are added, to a concentration of 0.1 to 0.5% yeast by weight of the fermenting broth. The fermentation requires 16 to 20 hours. At the completion of fermentation, as judged by disappearance of sugar,

the fermenter contents are centrifuged. The beer flows to a beer well, in preparation for distillation. The yeast cream is recycled to the next fermentation. Recycling yeast has a number of advantages: the yeast becomes acclimatized to the peculiarities of wood sugar solutions, fermentation is more rapid, and fermentation of galactose, which is scarcely used in this time frame by fresh yeast, is improved.

For distillation, the beer is pumped to the stripping section of the distillation columnn. Alcohol is boiled off the beer and passes to the rectification section of the column, where its concentration is raised to, or near, the azeotropic concentration, 95% by volume alcohol. The pilot plant is not equipped to raise the concentration beyond 95%, to the 99+% level required for use as a gasoline extender.

From the base of the column the stripped liquor, called stillage, is collected for analysis and study. One option for use of this effluent is to convert its sugar content to single cell protein. A small plant to accomplish that is installed, and its general structure, which is quite conventional, is outlined on the right hand panel of the control panel shown in Figure 5. It is felt by FRI scientists that this study is complete and other approaches to effluent utilization are under investigation, but are not a part of pilot plant operation.

3.4 Results

3.4.1 Radiata Pine

Alcohol Yields

The yield of alcohol was 205 kg per tonne of pine dry substance, or 260 litres per tonne, on an anhydrous basis. The pilot plant produces only 95% alcohol, and the yield of this product would be 5% more, or 274 litres per tonne. The "efficiency" of the process needs to take into account the chemical composition of radiata pine, and the yields of the various sugars, fermentable and non-fermentable. The composition of the pine, on an extractive-free basis as given in (2), follows:

Radiata Pine Composition
(dry, extractive-free)

	%	
Carbohydrate		
glucan	43.2	
mannan	9.3	
galactan	4.0	
xylan	7.3	
arabinan	2.8	66.7
Lignin		26.5
Other (ash, O-acetyl, 4-MeO glucuronic anhydride, acid-soluble lignin)		6.8
		100.0

If only glucose and mannose are taken to be fermentable sugars (some slow fermentation of galactose adds little to alcohol production) the theoretical yield of fermentable sugars on this basis is 57.8% of the dry wood. Successful fermentation of galactose and the pentoses xylose and arabinose would increase the theoretical yield to 72.3%, but this is still not available. Also these figures need some correction for the extractives content, which is highly variable in pine, and averages about 5% of the dry wood. On a dry pine wood basis, then, the theoretical yield of fermentable sugars is 54.9%, or of total sugars 68.6%.

Yields of all sugars varied between 48 and 52% of the weight of the wood, and yields of fermentable sugars (glucose + mannose) were 33.4%. This latter represents a hydrolysis efficiency of 60.8%. Total sugar production efficiency was 72.8%.

FRI calculations are quite close to these figures, giving cellulose hydrolysis efficiency as 62% and hemicellulose hydrolysis efficiency as 76%.

These yields are as good as those reported for acid hydrolysis operations by anyone, and indeed better than most, much better than commercial operations as reported, which run in the 45 to 55% range.

Sugar Concentration

Sugar production of 500 kg/tonne of wood is accomplished by a percolation procedure described above, the volume of which is 20 m^3 of percolate, resulting in an average sugar concentration of 2.5%. Of this sugar, 81.4% is fermentable, the concentration of fermentable sugars being 2.0%. This is of course quite dilute, since fermentation can produce only 1% alcohol to be distilled. This is not impossible, since such concentrations are also characteristic of sulphite alcohol plants around the world.

As will be discussed later, within limits sugar yield and concentration are inversely related in acid percolation systems: higher concentrations are achieved at lower yields. Plans for increasing sugar concentrations by recycling percolate will also be discussed.

Total Mass Flow

Mackie of FRI has pointed out (2) that such yields represent a substantial degradation of carbohydrate, about 24% of the weight of the wood. Carboydrate degradation products affect the processability of the resulting hydrolysates, as well as the smooth operation of the plant. A total "mass flow", from (1), is shown here. It shows essentially the following yields: ethanol 20.5%, carbon dioxide 21.5%, lignin 26.8% and stillage 36.7%. Within the stillage, 7% is pentose sugars (plus some galactose), while 29%

are tars, both as solids and as dissolved "humic substances". Stillage components to which much attention has been directed in the literature, such as furfural, hydroxymethylfurfural, laevulinic acid and formic acid, together amount to 4.65%.

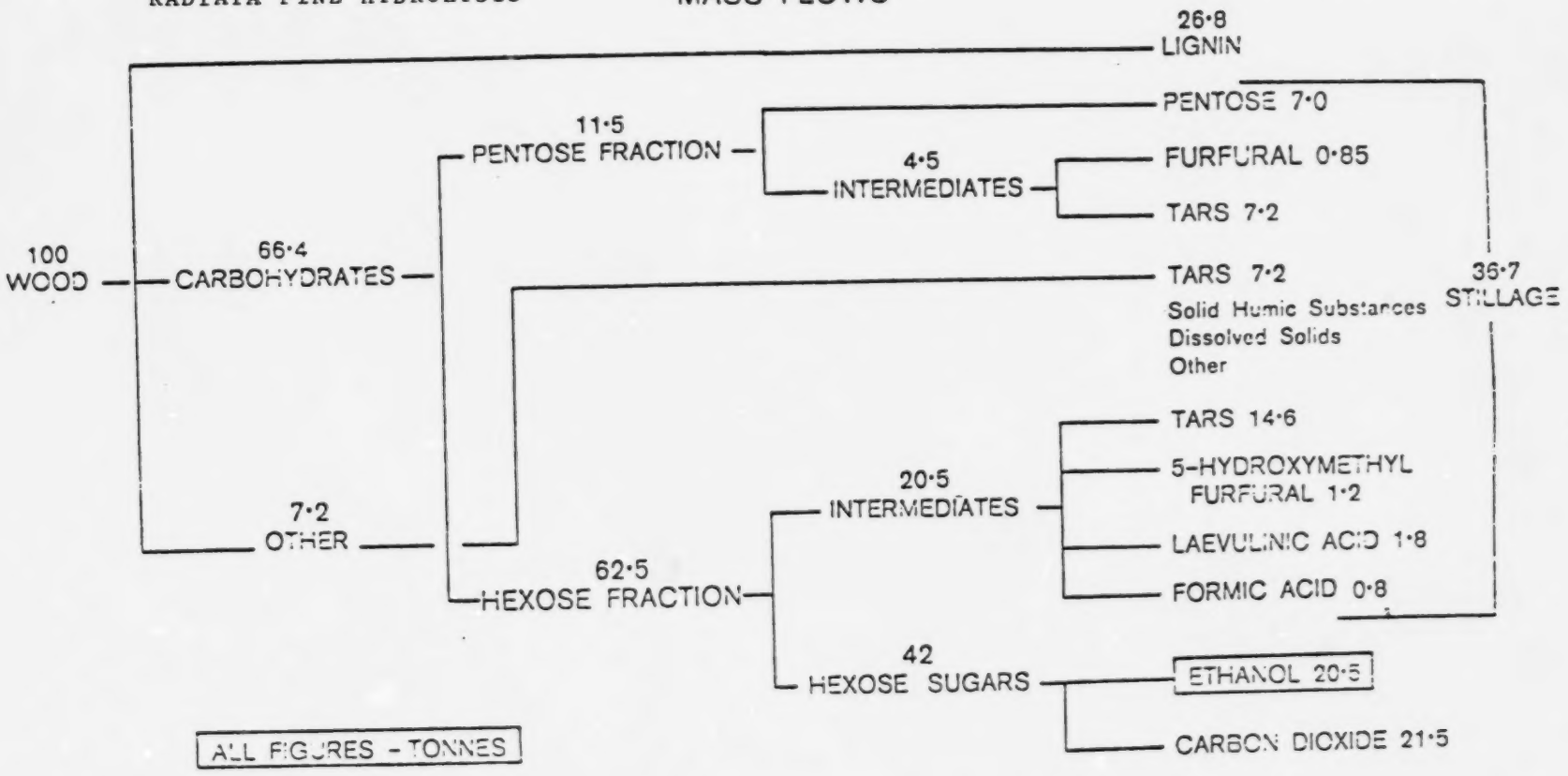
The mass flows shown lump together everything but hexose sugars and lignin in the "stillage". This is incorrect since volatiles are collected. However, these are of small amount, only about 1%, containing furfural, methanol and 2-butanone. They are not considered of sufficient quantity to be of economic interest except in a very large plant.

A great deal of attention has been directed in the past at the potential toxicity of the furfurals, and at the importance of utilization of the pentoses. Mackie (2) has pointed out that the real source of fermentation inhibition is more likely to be the humic substances, since the yeasts are quite tolerant of furfurals at these concentrations; and that while pentose fermentation is very important from a yield point of view - and this would be especially true for poplar - successful fermentation or protein production utilizing pentose is not a solution to stillage as an effluent pollution problem. Stillage has a COD 23,000 mg per litre and BOD 14,000 mg per litre and is highly coloured.

The recovery of lignin is essentially quantitative, although the lignin product has

RADIATA PINE HYDROLYSIS

MASS FLOWS



been highly polymerized in the process, and has reacted with and incorporated some of the carbohydrate degradation products. It has, however, retained its physical structure to a large degree and presents no serious handling problem. It has a high calorific value, 25.8 MJ per kg (11,100 Btu per lb.), and would be suitable for combustion after pressing in something like a bark press.

Energy Balance

Radiata pine has a calorific value of 20 MJ per kg on a dry basis. As received, at 53% moisture, the net calorific value would be 17.45 MJ per kg, or 17,450 MJ per tonne.

The heat value of the products may be calculated as follows:

	MJ/kg	Qty		MJ
Ethanol	- 26.7 x	205 kg	=	5,475
Lignin	- 25.8 x	268 kg	=	6,920
Stillage, via methane	- 50.3 x	90 kg	=	4,525
				16,920

The production of methane from stillage is discussed in (2), and the figure given appears to be reasonable. Heat recovery from wet lignin may be about 4,200 MJ/tonne pine. Total energy recovery in the products would be 71% of the energy of the wood.

This calculation, which shows a rough balance between the heat contents of the wood and the products, ignores processing energy for hydrolysis and ethanol distillation.

3.4.2 Poplar

At the time of undertaking this report, some batches of poplar, both from New Zealand and from Canada, had been processed. Results were available on the New Zealand poplar, designated Populus tremula, presumably very close to Canadian aspen P.tremuloides. Very recently the results of processing Canadian aspen have been made available to us by Canertech.

Alcohol Yields

Alcohol yields with poplar averaged 185 litres per tonne, substantially less than obtained with pine (260 litres per tonne). This is partly due to the higher pentose sugar content of the poplar, the composition of which is as follows:

	Populus Tremuloides Composition (dry, extractive-free) %	
Carbohydrate		
glucan	50.4	
mannan	2.2	
galactan	0.6	
xylan	19.4	
arabinan	0.4	73.0
Lignin		19.0
Other		8.0
		100.0

Theoretical yields of fermentable sugars on this basis are 57.6%, or based on dry wood (including extractives of 3.9%) 55.4%. Total sugar theoretical yield is 78.3%.

Yields of fermentable sugars were 31.0% of the wood, representing 53.8% hydrolysis efficiency. Yields of total sugars were 46.6%, representing 59.5% conversion efficiency. These conversion efficiencies are decidedly lower than those obtained with pine (60.8% versus 53.8%; 72.8% versus 59.5%). These low conversion efficiencies result in the low alcohol yield, 185 litre/tonne wood. It may be that further work will show some improvement but unless xylose fermentation becomes feasible alcohol yields will remain well below those for pine. An economic analysis based on these results appears in Section 5.3.

Lignin

Another factor adversely affected these results: the physical form of the lignin during and after hydrolysis. It collapsed to form a fine particulate mud which interfered with percolation and made it very difficult to remove solution from the reactor. For this reason, percolation was stopped when 7 to 12% of the total available carbohydrate remained in the residue, unhydrolyzed. To avoid this, coniferous wood, would have to be mixed with poplar to some undetermined extent, probably 25%. This requires further study in the pilot plant.

Energy Balance

While no studies are reported on stillage from poplar, it would be expected to be similar to that from pine, except that since there are lower yields, more stillage can be expected. Also there is less lignin by about one-third, so an adverse energy balance can be expected.

3.5 Discussion of the Technology

This discussion, at this point, is concerned only with the technology, and not with the economics, insofar as these are separable. It will deal with the strengths of the process, and its weaknesses in terms of those which can be corrected and those which are inherent.

3.5.1 Strengths

Experience

The FRI pilot plant is based upon the only process now operating industrially. There are about 40 plants in the USSR which are in operation, some of them for the last 30 years, in which wood wastes from forest industries are hydrolyzed to fermentable sugars, which are then either fermented to alcohol or to single cell protein for cattle feed. The results of their experience are generally available (24). Recently the Brazilian company COALBRA adopted this process for its wood-to-ethanol projects. These are now under construction and have the dual purpose of making charcoal for the steel mills (CO in the company name) and alcohol (AL), and therefore incomplete hydrolysis is not of basic concern as long as the lignin is recovered for the charcoal.

The pilot plant itself has tried many operating variations, and these and their other experiences have been very openly described (1, 2, 6).

A plant based upon this process is now under construction at Rainbow Springs, about 50 km from the pilot plant, and is expected to be in operation by early in 1983. The experience of this plant will also be available to interested parties.

Yields

The alcohol yields from pine are excellent for an acid hydrolysis process, and the hydrolysis efficiency is better than has been reported by others. The yields and efficiency from poplar are not as good, indeed are so low as to cause concern.

Wood Preparation

This process accepts chips or sawdust in conventional form, and no wood preparation is necessary. The presence of bark on the wood is tolerable up to quite high levels, although bark occupies reactor space without contributing to sugar production.

Once the chips are in the digester, they are not handled further, but are treated in place. Other acid hydrolysis processes of two or more stages have to remove the chips from the reactor after the first stage and separate the sugar solution from the lignocellulosic fibre. This is very difficult.

Overall Design and Potential for Improvement

The overall design is relatively simple compared to other proposed processes. It has many design weaknesses in its present form, and there is considerable potential for improvement, which will be discussed in the next section.

Several design details are good. One of these is the teflon mixing nozzle for sulphuric acid with water at the inlet to the reactor.

3.5.2 Weaknesses which can be Corrected

Experience

Although designed for continual operation, personnel assignment permits only about two batches per week to be run, the rest of the time being spent on analysis, tests and related experiments. In theory, in a 4 or 4 1/2 hour cycle, five or six batches can be run in 24 hours, or 25 to 42 batches per week, thus processing about one tonne of wood per week, to make about 260 litres of alcohol. The pilot plant program does not provide for this level of operation, so some problems, such as tarry precipitates interfering with reactor operation or hindering filtration, which are merely a nuisance on intermittent operation may become a serious limitation to production efficiency when operating every day and around the clock.

Corrosion

The metal used in construction of the reactor is Incoloy 825. It is better than the stainless steels which have been tried, but it is not satisfactory. Chromium as an alloying material is believed to catalyze sugar degradation, which eliminates all of the common stainless steels.

The Russians (and the Germans before them) used ceramic-lined reactors, with acid-proof brick, but it is reported that they have experienced high maintenance costs. They are now trying titanium-based alloys. FRI speculates that fluoropolymer or glass lining would be satisfactory, and will try fluoropolymer.

The experience of others in handling hot, dilute sulphuric acid has also been unsatisfactory. However, it is probable that zirconium or tantalum would perform satisfactorily. These are expensive, but perhaps less so than the present materials in use.

An alternative which may be even more satisfactory is to abandon sulphuric acid as the source of hydrolyzing catalyst. Hydrochloric acid is far more cost effective and corrosion by hot dilute hydrochloric acid can be effectively resisted or stopped by alloys now available. These concepts are discussed further in the next section.

Neutralization, Sulphiting and Heat Economy

As the FRI pilot plant is operated, hydrolysate from the first flash tank, which is at 130 to 140°C, is cooled to 70 to 90°C in order that it may be neutralized in open vessels. The neutralized hydrolysate, after some settling, is then heated to 130°C for the sulphiting treatment. It is then cooled to 90°C for filtration. This alternating of heating and

cooling is very wasteful of heat, and the relatively cool filtration leaves more $CaSO_4$ in solution than is desirable. It was not intended that the pilot plant be operated in this fashion. In the original installation, an in-line mixer was provided following the flash tank in which neutralization was to be carried out, thereby maintaining the temperature for the sulphiting step and perhaps also for filtration. Unfortunately, the in-line mixer did not function well and was abandoned, resulting in greater heat consumption.

It is not clear whether high temperature filtration was provided in the original plant design, although $CaSO_4$ is much less soluble at 130 to 140°C than it is at 90°C, thereby leaving less inorganic salt in solution. In any rebuild of the pilot plant, these factors will be taken into account. However, they will not cure the whole problem, since the $CaSO_4$ is contaminated with humic substances and tars, and is very difficult to filter. Some combination of settling and filtration will have to be employed.

Acid Distribution in the Reactor

When operating in a percolation mode, that is when the tower contains only wet wood and acid solution is percolating down through it, the distribution of acid within the bed of wood is a matter of some concern. FRI has apparently avoided this problem by keeping the bed of chips

covered with acid at all times, with equal rates of drainage and acid addition. This approach is costly in terms of sugar concentration. The Russians, and the Swiss, have approached this problem differently, by providing acid distribution systems within the chip bed, and an acid collection system nearby. This is a matter of refinement of design the direction of which is reasonably clear.

Fermentability

The sulphiting treatment has improved the rate of fermentation, reduced the amount of yeast required, and perhaps increased alcohol yields. It is far from perfect; the colour of the hydrolysate is imparted to the yeast, and generally there is a belief that this is an area where further research is warranted.

Effluent Treatment

The stillage from this plant is heavily loaded with BOD and colour (2). An approach considered valuable in alleviating this condition is anaerobic digestion to methane. Anaerobic digestion is said to remove 90% of the stillage organic matter while producing 90 kg methane per tonne of wood. As pointed out later (p.55) such an installation would add $1.7 million to the capital cost of a 10 million litres/year ethanol plant, and would be accompanied by savings of 5.9 cents/litre in energy costs.

Poplar Hydrolysis

The alcohol yields from poplar in this pilot plant are quite disappointing. It is a moot question whether manipulation of operating variables will produce any significant improvement. The only hope for substantial improvement would be the successful fermentation of xylose. This would raise alcohol yield from 185 litres per tonne level to 260 to 280 litres per tonne, which could place it more or less on a par with the yield from pine.

Current opinion is that based on the research efforts at NRC, at Purdue and at Peoria, a viable process for xylose fermentation is imminent. This opinion has been strongly held for about two years. The research continues to result in improvements which should solve this problem within the next two or three years.

3.5.3 Weaknesses Inherent in the FRI Process

Aside from its suitability for poplar, as discussed above, some problems are inherent in the system upon which this pilot plant is based.

Sugar Degradation and Humic Substances

The operating problems associated with the formation of humic substances due to sugar degradation, and also due to reaction of lignin with sugar degradation products, are described at some length by Mackie (2). About one-third of

the wood fed to the reactor ends up as tars and dissolved humic substances. These coat vessels and pipes, and also contaminate the $CaSO_4$ from the neutralization of the hydrolysates. They contribute heavily to the effluent disposal (or utilization) problem. While other operators have not stressed this aspect of the process as strongly as has Mackie, it appears to be inherent in a sulphuric acid high temperature wood hydrolysis process. Some efforts have been made to identify substances which may lessen sugar degradation, such as $FeSO_4$, but these experiments have not been systematic, they have not covered a wide range of substances and generally one feels that the FRI operators are not optimistic about this approach. It is a serious matter, requiring attention.

Scaling-up to Full Plant Design

Any process based upon percolation has constraints upon the geometry of the system. These constraints arise in part from the time required for acid to drain from top to bottom of the bed of chips, which sets an upper limit to the depth of that bed; and from problems of acid distribution within the bed, which may limit its diameter. Longer percolation time results in a lower temperature reaction, and low temperatures are associated with low yields. Ideally one should aim for higher temperatures. Mackie (2) says there are practical upper limits to the temperature at which conventional percolation systems may be operated as at higher temperatures

liquor residence times in the reactor must be correspondingly shorter if yield benefits are to be realized. This limit is around 200°C and hence for temperatures higher than this other hydrolysis systems, e.g., plug flow or extrusion, must be considered.

The twin reactors in the Rainbow Falls plant are each 10 m in height. This may be approaching the upper limit of height. This plant will make six million litres per year of ethanol, which is not a large plant. To increase capacity, it may be necessary to treat the Rainbow Falls plant as a module and multiply it for larger output. This is expensive; and we do not yet know how well it will work.

Lignin Quality

The lignin produced by the FRI process is highly polymerized and otherwise modified. It has a good calorific value, and is useful as fuel after pH adjustment and pressing. However, its other potential uses are probably severely limited, compared to autohydrolysis or enzymatic lignins.

High value lignin markets have thus far developed only to very limited tonnage (vanillin and a few other chemicals). This is an area of active research. When such markets do develop, this acid treated lignin may not qualify.

Batch versus Continuous

This is inherently a batch process, with all the associated disadvantages. These include low capacity for a given volume and the associated high capital cost. There is probably also more variability from batch to batch than in a well controlled continuous process, as the Canadian pulping industry can testify. Continuous processes are generally preferred by engineers.

3.5.4 A Second Generation Pilot Plant

The FRI scientists are well aware that the existing pilot plant is far from what they would like, indeed it does not even match their original concept.

Main Hydrolysate Recycle

There are two hydrolysates, the prehydrolysate produced at 150°C and the main hydrolysate produced at 185°C. Both of these contain sugars, sugar degradation products, acid and heat. It has been stated by FRI scientists that sugar degradation in the presence of sulphuric acid is minimal at 150°C. Their concept, therefore, is to cool the main hydrolysate, perhaps by flashing from 185°C to 150°C, and to use it for prehydrolysis. This would save heat and acid, and would have the further great advantage that sugar concentrations would be raised from about 2 to 2.5% up to 4 to 5%. This would benefit fermentation and distillation.

FRI are now buying a second reactor. This will be teflon-lined, probably by a Japanese supplier, to avoid corrosion and reduce metal-catalyzed sugar degradation.

The intention is to operate the two reactors with recycle of main hydrolysate; wood chips in Reactor 2 would be prehydrolyzed with main hydrolysate from Reactor 1. The prehydrolysate, which now will contain the hexose sugars from Reactor 1, will be neutralized, sulphited, filtered and fermented. Then fresh acid will be used to carry out the main hydrolysis in Reactor 2, and the hydrolysate so produced will be used for prehydrolysis in Reactor 1. In stable operation, the two reactors will be operated alternately, and out of phase, fresh wood going to one reactor while fresh acid goes to the other, and the main hydrolysate from one reactor being used for prehydrolysis in the other. Instead of main hydrolysate being processed for fermentation, prehydrolysate containing recycle sugars from a completed main hydrolysis in the other reactor will proceed to further processing and fermentation. It can be considered a counter-current flow of acid and wood.

The savings are expected to be substantial. A word of caution is necessary, however; sugars are not immune to degradation by sulphuric acid at 150°C, the extent of degradation to be expected being between 5 and 15%. However, by keeping the time of reaction as short as possible, the lower of these two figures may be realized.

In-Line Neutralization and Temperature Smoothing

The treatment of the hydrolysate ahead of fermentation involves neutralization, sulphiting and filtration. In-line mixers are available which should function quite satisfactorily to enable neutralization with lime to be carried out at hydrolysate temperature, at which temperature $CaSO_4$ is much less soluble than it is at 90°C, and much more complete removal is possible. With a pressure vessel for sulphiting, as now, and a pressure filter, the three operations may be carried out in sequence without ever lowering temperature. Thus no cooling and heating are required until the sequence is completed. At that point, there is enough heat in the sugar solution to warrant some heat recovery.

In Section 4.0 of this report, capital cost calculations are presented for a full scale alcohol plant based upon this improved sequence, and a flow sheet of this second generation plant, shown in Appendix 1, accompanies the calculation.

4.0 ECONOMIC ANALYSIS

4.1 Investment

Whitworth (5) has provided the following capital cost estimation for a 30,000 litres per day, 95% ethanol plant based on radiata pine:

Capital investment for a 30,000 litres ethanol per day plant based on pine

1.	Capital Equipment		
	Raw material preparation	85,000	
	Hydrolysis	1,358,000	
	Neutralization & treatment	289,000	
	Fermentation	177,000	
	Distillation	1,100,000	
	Effluent	171,000	
	Hot water supply	565,000	
	Site development	200,000	
	Buildings, facilities	621,000	
	Other	252,000	4,818,000
2.	Contingency (10% capital)		481,000
3.	Engineering charges		241,000
4.	Commissioning charges		120,000
5.	Working capital		294,000
			$5,996,000

Thus the capital cost is very close to $6 million for 9.9 million litres per annum, or 60.6 cents per annual litre (when FRI made this estimate, the N.Z. and

Canadian dollars were very close to equal). A dehydrating unit would increase this cost by 15% and decrease output by 6%, resulting in a capital investment of 73.4 cents per annual litre for a plant for anhydrous ethanol to make 9.3 million litres per year (28,200 litres per day).

The economic analysis by Whitworth, just presented, is based on the FRI plant design. It has been impossible for us to determine, from their publications or from interviews, the detailed basis of their estimates. We have made our own economic analysis based upon the "second generation" plant design: two brick-lined reactors operating out of phase, main hydrolysate recycle, in-line neutralization, sulphiting and filtering. The flow sheet which forms the basis of our calculations is shown in Appendix 1, Process Flow Sheet 2. This plant includes a dehydrating unit to produce anhydrous ethanol. It also includes facilities for pressing the lignin and feeding it to a boiler. It is based upon Canadian costs of equipment and construction and is expressed in 1982 Canadian dollars.

Investment for a 30,000 litres per day anhydrous ethanol plant based on pine, improved design; 330 days per year, 24 hours per day; 9,900,000 litres per year, 100% ethanol.

	$000
1. Capital Equipment	
Raw material storage and distribution	320
Hydrolysis	2,300
Fermentation and yeast recovery	1,250
Distillation and dehydration	1,010
Alcohol product storage	140
Boilers and utilities	1,910
Boiler house and process buildings	930
Site development	150
Office, personnel and maintenance facilities	80
	8,090
2. Engineering and project management	1,330
3. Interest during construction	810
4. Working capital	420
Total investment	10,650

This estimate, which does not include land or infrastructure costs, is $1.08 per annual litre of anhydrous ethanol, 77.6% higher than FRI's estimate, which was for 95% ethanol and therefore for a 6% larger volume. The extra costs are due to a two-year time gap in estimating, to the inclusion of alcohol dehydration facilities, a more realistic cost estimate for hydrolysis reactors, much higher utilities facilities costs due especially to the high cost of a lignin-fired boiler, and higher building costs due to climate.

Climate considerations have also led to higher working capital, which includes feedstock inventory. In addition, this investment estimate includes interest during construction, which was omitted from the New Zealand estimate. The present estimate is believed to be realistic for a plant to be built in Canada now.

This investment calculation has omitted provision for effluent treatment. The manner in which effluent will be treated is highly site specific, several alternatives being available, including anaerobic fermentation to generate methane. The latter installation would add about $1.7 million to the capital cost and save about $590,000 in energy cost. This can well be a later stage in plant development. The lignin boiler contributes about $1.5 million to capital cost and this could be saved by buying steam from an existing facility. This may prove to be the case where a wood-based alcohol plant is built alongside an existing pulp mill or other steam-generating operation. However, failure to make use of the heat content of the lignin could add to production costs by as much as $500,000 per year, depending upon the cost of steam.

4.2 Production Costs

Whitworth (5) has provided operating cost data for a 30,000 litres per day of 95% ethanol plant based on radiata pine. Whitworth's data are compared with our calculations in the following table. It should be recognized that our calculations have a different basis, being based upon the improved plant design discussed above and anhydrous ethanol.

Production costs for ethanol plant:

Production costs for ethanol plant
based on pine, 30,000 litres per day

	Annual Costs, 1982 Cdn. $000	
	Whitworth (5) 9.9×10^6l 95% EtOH	This Report 9.9×10^6l 100% EtOH
1. Raw material		
wood @ $45/tonne	1,680	1,780
sulphuric acid		
@ $96/tonne	190	
@ $310/tonne		300
lime @ $82/tonne	160	80
sod. sulphite		
@ $452/tonne	90	90
sod. phosphate, urea molasses	60	60
	2,180	2,310
2. Energy	70	540
3. Labour and management	600	940
4. Repair and maintenance	120	520
5. Overhead, insurance	140	140
Total, annual	$3,110	$4,450
per litre	31.4 cents	44.9 cents

The present estimate is 43.0% higher than the FRI estimate. The differences are due to the much higher provision for energy costs in our estimate, as well as higher labour and management costs, and higher repair and maintenance costs. It is believed that these production costs are realistic for a plant of this design to be built in Canada now. These production costs do not include energy taxes, product distribution,

depreciation or profit, which are discussed further on in this report.

Wood cost represents 54% and 40% of production cost in the two cases. Obviously the cost per litre is very sensitive to the cost of wood, as discussed below.

4.3 Profitability and Sensitivity

Profitability depends upon the selling price at the plant gate. Whitworth (5) added 20% return on investment to his production cost and thereby derived an ex-works selling price of 47.4 cents per litre, a price which included 4.3 cents attributable to depreciation. If we add 20% return on investment, we calculate a plant gate selling price of 66.2 cents per litre.

Sensitivity

It is apparent that profitability of this enterprise will depend upon selling price, wood costs and capital costs. Wood costs of $45 per tonne, used by Whitworth and ourselves, are generous if the plant is to be based on wood residue. A more realistic price would be $30 per tonne and, if the plant is situated favourably, this may be even as low as $15 per tonne. Capital costs can be reduced by taking advantage of regional development and other government grants and subsidies. The return on investment of 20% of production cost before taxes is unrealistically low in the light of risks involved, and higher returns, such as 25% or 30% may well be required to interest private sector investors. On the basis of these parameters, we have calculated a plant gate selling price for three different wood costs per

tonne, $15, $30 and $45; for three different capital investment levels, $10,650,000 as calculated above, $9,650,000 which considers that $1 million in grants and subsidies can be assembled for the plant; and $9,150,000 which figure arises by omission of the lignin boiler and other steam generation; and for return on investment of 20%, 25% and 30%. All of these calculations are based on the newer plant design. The effect of effluent treatment on capital and operating costs as given on page 55 is a trade-off with a 2.9 year payback.

Recognizing these parameters as the most important gives rise to this equation:

$$P = 27.0 + 0.40W + \frac{RI}{A}$$

Where P is the plant gate selling price in cents per litre,
W is the cost of wood in dollars per tonne dry weight,
R is the required return on investment, as a percentage,
I is the investment in dollars, and
A is the alcohol production in litres per year.

At the base case investment of $10,650,000, the plant gate selling price P is:

Wood cost		$15	$30	$45
Return	20%	54.2	60.2	66.2
	25%	59.6	65.6	71.6
	30%	65.0	71.0	77.0

With capital reduced by subsidy to $9,650,000, the plant gate selling price P is:

Wood cost		$15	$30	$45
Return	20%	52.7	58.5	64.5
	25%	57.4	63.4	69.4
	30%	62.2	68.2	74.2

With capital reduced by omission of the steam generation plant to $9,150,000, the plant gate selling price P is:

Wood cost		$15	$30	$45
Return	20%	51.5	57.5	63.5
	25%	56.1	62.1	68.1
	30%	60.7	66.7	72.7

It is apparent from these figures that reduction in capital cost by up to 15% has only a small effect on the plant gate selling price of alcohol, 3 to 4 cents per litre. Wood cost has a significent effect, 2 cents per litre for each $5.00 in wood cost per dry tonne. Increase in return on investment from 20% to 30% has a larger effect on alcohol cost, 9 to 10 cents per litre.

The lowest selling price at the plant gate derived in these calculations is 51.5 cents per litre with wood at $15 per dry tonne and 20% return on investment of $9,150,000. More expensive wood and the higher investment, both part of the base case, result in a plant gate selling price of 66.2 cents per litre at 20% return on investment, or 77.0 cents per litre at 30% return on investment. Since these plant gate prices are

much higher than the refinery price of gasoline, they must be considered unrealistic, and the plant unprofitable.

Effect of Plant Capacity

In this report we have taken the plant capacity to be about 30,000 litres per day, because that was the basis for the FRI calculations. However, that is rather small for a viable alcohol plant, a more common economic size today being 120,000 litres per day, with larger plants also built recently and under construction. In the following table are compared investment and production costs for plants of this design to make 120,000 litres per day and 240,000 litres per day 100% ethanol.

Plant capacity litres per day

	30,000	120,000	240,000
Investment $000	10,650	25,500	39,500
Investment/litre, $	1.08	0.64	0.50
Plant gate selling price, cents/litre, at 20% return on investment			
Wood at $45/tonne	66.2	54.4	49.8
Wood at $30/tonne	60.2	48.5	43.8
Wood at $15/tonne	54.2	42.7	37.9

It is apparent that a plant to make 240,000 litres per day can sell anhydrous ethanol at 50 cents per litre with wood at $45 per dry tonne including 20% return on investment; a plant of 120,000 litres per day capacity requires wood at $30 per day tonne for the same plant

gate selling price.

In the accompanying graph, investment and plant production costs before return on investment are shown for these three plant capacities. At a capacity of 120,000 litres per day or more and wood costs of $15 per dry tonne, plant production costs below 25 cents per litre can be achieved, before allowing for return on investment.

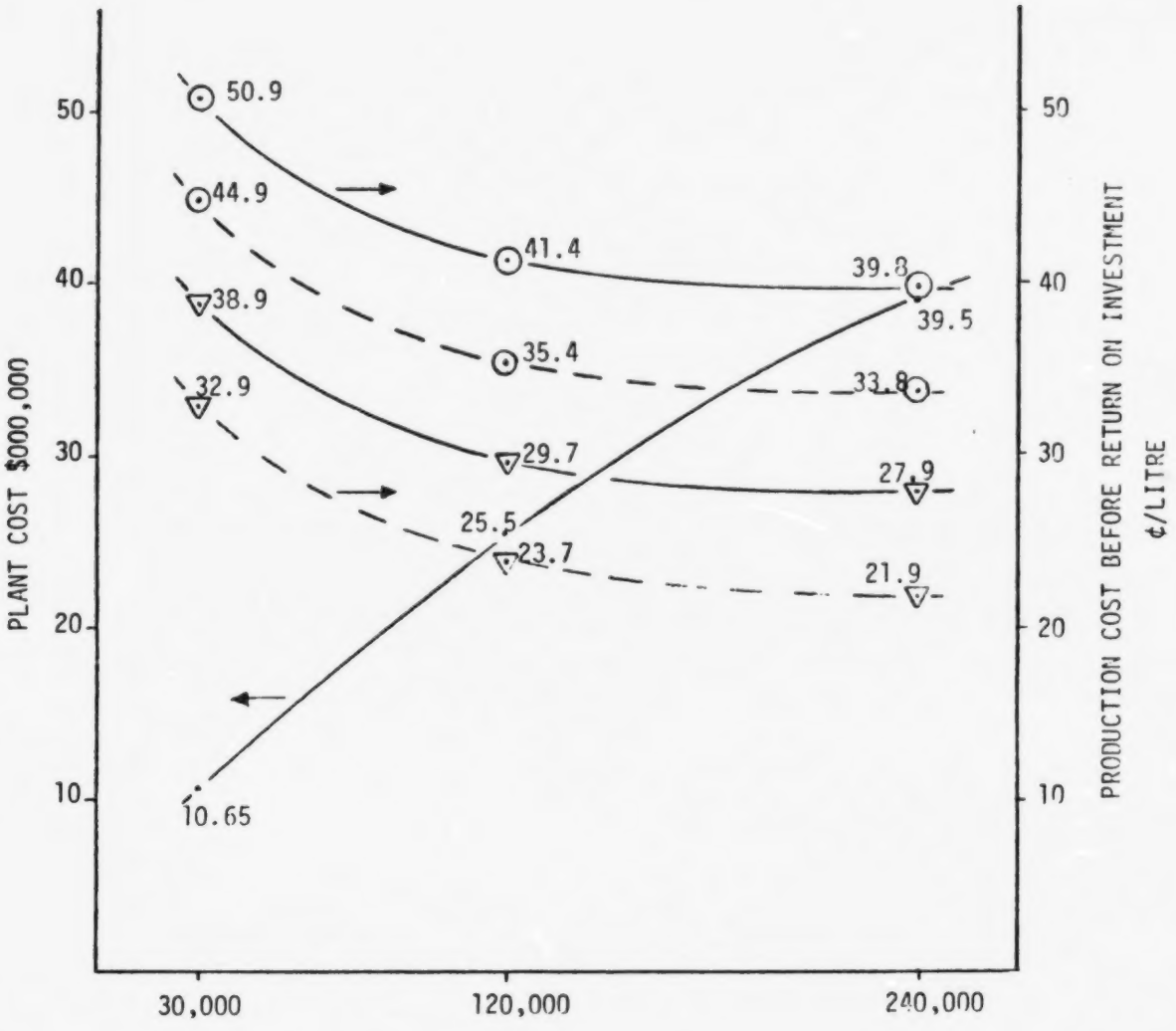
Profitability depends upon income. It cannot be calculated for these plants without knowing what income to expect. This question will now be examined.

Expected Income

Regular gasoline in Ontario has a pump price of 45.3 cents per litre made up, according to Shell Oil as quoted by the Toronto Globe and Mail on October 21, 1982, as follows:

	Cents per litre
Crude oil costs	7.6
Refinery margin	7.6
Crude oil royalties	10.6
Dealer margin	3.0
Federal tax	9.3
Ontario tax	7.2
	45.3

PLANT CAPITAL AND PRODUCTION COSTS



Based on this breakdown, if ethanol produced from wood is not subject to crude oil royalties, it would have to sell for 25.8 cents at the plant gate to compete with gasoline, including return on investment. This cannot be achieved at any reasonable return on investment in any of the cases discussed.

If we use the world price of crude oil, at $34 per barrel of 164 litres, as a basis, and allowing for 20% loss in value in refining to gasoline, we can calculate the cost of gasoline at the refinery gate, including refineries margin, at 32.5 cents per litre. No 30,000 litres per day FRI process plant can meet that plant gate price. A larger plant, making 120,000 litres per day or 39.6 million litres per year, can expect an income of $12,870,000. With wood at $15 per dry tonne, production costs would be $9,380,000, and return on investment of $27,300,000 would be 12.8%. This would not be considered a satisfactory profitability by private investors today.

The above calculation was made on the basis of the world price of crude oil. We do not have real production costs of tar sands oil, Arctic oil or Cold Lake heavy oil. If these are higher than the world crude oil price, wood-based ethanol plants may be equally profitable.

Other Factors Affecting Profitability

In this discussion of profitability, no allowance was made for income derived from by-product sales, such as lignin, nor was any credit given to alcohol production from fermentation of pentose sugars. These can have a

large impact on profitability as can government policy on taxation or upward movement in the pump price of gasoline. These are discussed in Section 5.0.

5.0 POTENTIAL FOR COMMERCIAL DEVELOPMENT

In this section we shall discuss the potential for commercial development of the FRI process. As indicators of potential, we shall employ primarily economic criteria as they apply to Canada. This section also includes a comparison with other processes, and comments on the availability of cheap feedstocks.

5.1 Requirements for Economic Acceptability

For ethanol produced by the FRI process to become economically competitive with gasoline would require one or more of these changes:

(i) Favourable taxation policy
(ii) Sale of lignin
(iii) Pentose fermentation
(iv) Gasoline price increase
(v) Decreased feedstock costs
(vi) Use as an octane enhancer
(vii) Reduced sugar degradation in process

These can now be considered in the above order.

(i) Favourable Taxation Policy

It is difficult to believe that major tax concessions can be made when federal and provincial fiscal responsibilities and deficits are so high. However, it is worth considering what degree of concession would be required to achieve 20% return on investment. A pump price including this return for alcohol made by the FRI

process in the base case would have these components:

	Cents per litre
Taxes	16.5
Production cost	44.9
20% return on investment	21.3
Dealer margin	3.0
	85.7

To bring the 85.7 cents pump price of alcohol to the 45.3 cents pump price of gasoline only requires the reduction of gasoline tax by 4.1 cents per litre for gasoline containing 10% alcohol. This is 25% of present combined federal and Ontario gasoline tax. Such a concession could be considered reasonable when the effects on employment and welfare are taken into account. It would serve to bring in new industry and new technology. A case can, perhaps, be made for it.

(ii) Sale of Lignin

In the present plant cost estimates, lignin is considered a fuel, with no outside sale. In the FRI process employing sulphuric acid hydrolysis, the lignin is highly polymerized and is unlikely to find higher value uses.

If a market could be found for lignin at 25 cents per lb., or 55 cents per kg, a figure which is heard from time to time, this would amount to 55 cents per litre of alcohol. If 20% of this were the cost of marketing, drying, packaging and delivering the lignin, and the rest were profit and applied to the cost of alcohol, the

pump price of alcohol fuel would then be 41.7 cents, below today's pump price for regular gasoline, 45.3 cents. This exercise explains the great effort to promote the sale of lignin at this price and in this volume. However, such markets have not developed. If they did, they would not apply to FRI lignin.

(iii) Pentose Fermentation

The possibility of fermenting pentose sugars can be expected with confidence to be available in two to three years. This will increase alcohol production by 13%, raising the output from 242 to 273 litres per tonne of pine. This would result in a saving of 5 to 6 cents per litre in the cost and the price of alcohol.

Such a development will be far more important with poplar than with pine, where the 40% increase in alcohol production will be large enough to materially affect the capital cost of the plant as well as production costs.

Pentose fermentation has a significant contribution to make to the viability of wood-based alcohol plants. It is unlikely to make the difference between economic acceptability and unacceptability for pine-based FRI process alcohol plants.

(iv) Increase in Pump Price of Gasoline

This report need not speculate on the future of gasoline prices, except perhaps to say that there is a gap of about 40.4 cents a litre between the present pump price of gasoline and our base case alcohol pump price. Any upward movement obviously improves the prospects for

alcohol for liquid fuel, but it would be unrealistic to expect this alone to result in viable pine-based FRI process alcohol production in the foreseeable future.

(v) Decreased Feedstock and Operating Costs

There is not much room for achieving great reductions in operating costs. However, about 2.5 cents per litre is the cost of sulphuric acid and the cost of its neutralization. By using a cheaper acid such as SO_2, or a more effective acid such as HCl, this cost may be cut by about one cent per litre.

The cost of feedstock used in this calculation is $45 per dry tonne, which would be a stable price for pine wood. There should be no difficulty in locating sawmill residues or other pine or spruce residues at $30 per dry tonne, and quite possibly at $15 per dry tonne.

Each $5 reduction in wood costs would lower alcohol production cost by 2 cents per litre. The maximum expectation is 6 to 12 cents per litre.

(vi) Alcohol as an Octane Enhancer

Alcohol is well established as an octane enhancer, and the use of alcohol at a 10% or even 20% level would enable refiners to modify their operations to improve their costs. This was shown in "Feasibility of the Gasohol Plant in Ontario" by John C. Clark, of Imperial Oil Limited, a 1980 research study undertaken at the Faculty of Management Studies, University of Toronto. Clark concluded that the "integration of an alcohol plant with the refinery permits sufficient economies

that the cost of ethanol for a satisfactory return exceeds crude oil prices by only 30%", that is, 27 cents; and Clark suggested that "utilization of by-product pure carbon dioxide from the fermentation process as a supplemental feedstock to an adjacent methanol plant permits an improvement in the economics of the ethanol investment".

Clark's study was not based on the FRI process, but it did serve to show that integration with an oil refinery could bring down the net cost of ethanol. Because of changes in the price of crude oil, and process differences, this cannot be quantified for the FRI process. It does, however, explain the continuing interest of various oil companies in alcohol technology development. With refineries operating at 80% or less of capacity, they are not rushing into this technology.

(vii) Reduced Sugar Degradation in Process

Present alcohol production is based on cellulose hydrolysis yields of 65%, and fermentation of hexoses only. It is conceivable that hydrolysis yield may be raised to 85% by stabilizing sugars against decomposition in the presence of acid at hydrolysis temperature. The mechanism for achieving this improvement is not yet demonstrated, so this must be regarded as speculation, at least in part, but it has some scientific foundation. By this means, the yields of alcohol would be increased by about 31%, thereby requiring a smaller plant for the same output, lowered production costs and larger income. Accepting higher production in the same plant would lower production costs by 14 cents per litre. This is obviously a

research direction worth pursuing.

Ranking of Requirements for Economic Acceptability

Of the above considerations, the greatest impact could come from a 4.1 cents per litre tax concession; lower wood costs and pentose fermentation have the potential of making a substantial improvement in costs amounting to 18 cents per litre. While this would not be enough to move the base case to a profitable range, it could improve a larger plant to an acceptable return. Upward movement in the pump price of gasoline would be helpful. Sale of lignin from the FRI process without modification is considered unlikely. The use of alcohol as an octane enhancer has considerable attraction for oil refineries, but in this respect the FRI process has no advantage over any other. Reduced sugar degradation in process is still unproved but is a promising area of research.

5.2 Comparison with Other Processes

The Russian version of the Scholler process has been in operation for about 35 years and is said now to be practiced in about 40 wood hydrolysis plants. It has undergone changes and improvements, as evidenced by articles in their journal "Hydrolysis and Wood Chemistry" (24). They, like FRI, have had difficulty with reactor corrosion. Originally lined with acid-proof brick, recently they have begun lining digesters with a titanium alloy. In the technology which they have provided to the Brazilians, this titanium alloy is included. It has been suggested that this is in part motivated by a desire to develop trade in titanium with Brazil but that can only happen if the

alloy proves to to be successful.

The internals of the Russian reactor are arranged so as to permit shorter percolation times, acid addition and hydrolysate removal tubes being provided within the reactor. This arrangement would overcome some of the concern about the effects of upscaling of reactor size.

A recent paper in the Russian journal describes preparations for a plant trial of hydrolysate recycle, as foreseen by FRI.

While we do not have all the evidence at hand, there is serious doubt that the Russians have achieved as high yields as the FRI. This may not be serious where the feedstock is sawmill wastes, as in Russia, or plantation eucalyptus, as in Brazil. The Russians may have come closer to solving the corrosion problem, and they may have a better configuration for adding and draining acid. We have no data on their capital costs and operating costs.

The Inventa design (23) is another variation on the Scholler process, in this case backed up by more than a decade of successful operation of an alcohol plant at Ems in Switzerland, during the forties and fifties, and having a functioning pilot plant there now. The process as operated in this plant was basically a Scholler process with a modified percolation schedule, and yields of about 52 U.S. gallons per ton (216 litres per tonne). The most recent information (1982) is that they now believe that they can build a continuous reactor, an upflow reactor in which wood is fed in at the bottom, passes upwards through a zone in which dilute sulphuric

acid is added at 140°C, whereupon the prehydrolysate is collected through a centrally located tube; and then further dilute sulphuric acid at 185°C is added through a series of ports at the side of the reactor, while hydrolysate is withdrawn through a centrally located perforated tube; and finally wash water is added and the lignin is collected with the washings through an overflow at the top. Our information is that this reactor has not yet been built; we see many problems in construction and operation, but no insurmountable problems. The successful construction and operation of such a reactor may represent a step forward in technology over the simple Russian process. As a continuous reactor, it would appeal more to those engineers who select process and equipment.

Any data which we have seen about the costs of the Inventa process are not encouraging. The capital cost is very high, and they hope that sale of lignin and other byproducts will contribute sufficient income to make a plant based on their process profitable. For reasons given above, this report feels that such a hope is premature at least.

In addition to these variations on the Scholler process, which is what the FRI process is, the only other dilute acid hydrolysis process currently being offered is the New York University process (Rugg, Werner-Pfleiderer, 22). In the NYU process, an extruder is used as the reactor. Sawdust or waste paper in finely divided form is fed to the extruder, and after a brief steaming to about 230°C to 240°C sulphuric acid is added, and the extruder then discharges. The hydrolysis time is limited to about 20 seconds. The pilot plant at NYU is

now about two years old, and a great deal of data has been amassed. Yields have been quite low, sugar conversion efficiency being about 45 to 50%. This equipment is quite expensive, and no plants of this design have yet been sold. A different extruder manufacturer, Wenger of Sabetha, Kansas, has sold a small extruder to Bio-Hol Developments pilot plant under construction in Toronto. This will be used for prehydrolysis, and perhaps for main hydrolysis also. It is a stated intention of Bio-Hol to market its technology once it has been developed. It cannot be considered a competitor to FRI as yet.

While many other dilute acid hydrolysis processes have been reported, no other process has been offered to the market. Any company about to build a wood based alcohol plant must take the FRI process seriously. It is likely that other, and perhaps better, processes will become available in the next year to two. At the recent Royal Society of Canada International Symposium on Ethanol from Biomass, Winnipeg, October 1982, Wayman proposed steam explosion prehydrolysis followed by plug flow acid hydrolysis as having the best scientific and practical basis (M. Wayman, "Fundamental Principles and Their Practical Consequences in the Conversion of Biomass to Ethanol", concluding remarks at the RSC Symposium) but no company has as yet offered this process.

An alternative to those mentioned is an enzyme hydrolysis process, such as those offered by Iotech, or by Katzen and Emert. There are great advantages to enzyme hydrolysis: very high yields mean that there are little if any tars and other sugar degradation products.

The lignin is unchanged, in contrast to lignin subjected to acid hydrolysis which is highly polymerized and therefore has a very restricted range of uses compared to enzyme lignin. The main difficulty with the enzyme processes is the very high cost of the complex enzyme systems required to effect cellulose hydrolysis. Should cheap enzymes become available, or should high value markets for lignin become available or a combination of both, then the enzyme process may become a strong competitor to the acid hydrolysis process.

5.3 Poplar Processing Economics

There is considerable interest in hydrolysis of poplar, especially aspen, in Canada because of its high mean annual increment, about 10 million tonnes, of which only about 10% is presently utilized for any purpose. In consequence, Canertech proposes to build a pilot plant and a demonstration plant for hydrolysis of aspen poplar, and in preparation has had a shipment of this wood processed in the FRI pilot plant. It has kindly made the results available to us. Yields of 185 litres of alcohol per tonne of wood (dry basis) are very low compared to those obtained with pine (260 litres of alcohol per tonne of wood, dry basis). This has been described and discussed on pages 37 and 38. Here we shall examine the economic consequences of these low yields.

Compared to a plant based on pine wood, one based on poplar would have 20% higher capital cost for the same output, that is for a capacity of 9.9 million litres/annum, the plant would cost $12,780,000, compared with $10,650,000 (p.54). Production costs are calculated to

be significantly higher, as follows:

Production costs for ethanol plant based on aspen poplar 30,000 litres per day, or 9.9 million litres per year.

		$ 000
1.	Raw material	
	wood @ $45/tonne	2,370
	chemicals	700
		3,070
2.	Energy	650
3.	Labour and management	1,035
4.	Repair and maintenance	600
5.	Overhead, insurance	155
	Total annual production cost	$5,510
	Production cost per litre	55.7 cents

Compared to 44.9 cents per litre price on the same basis (p.56).

Thus compared to pine wood feedstock, the FRI process applied to poplar has a production cost disadvantage of 10.8 cents/litre, as well as a capital disadvantage of 20%, or $2,130,000 for a plant designed to make 9.9 million litres per year. The major factor contributing to these higher costs is the high pentosan content of the wood. In the event that a practical process for fermentation of pentoses to ethanol becomes available, the yield of ethanol can be expected to increase by 85 litre/tonne of wood, that is to 270 litre/tonne. This

would change the economics markedly. The capital cost of the plant and the ethanol production cost would then be the same as for pine, with the addition of the extra capital cost associated with pentose fermentation equipment and the extra production costs associated with the operation of the pentose fermentation process, both presently unknown.

This is obviously a direction meriting considerable research effort for any group planning an ethanol plant based on poplar.

5.4 Availability of Cheap Feedstocks

At the recent Royal Society of Canada International Symposium on Ethanol from Biomass, mentioned above, R.D. Hayes of Agriculture Canada presented figures to show that there are now available in Canada 5 million tons of agricultural crop residues surplus to soil and animal needs, which can be delivered at low cost to an alcohol facility. At the same Symposium, Dr. M.R. Clarke of Forintek Canada presented data which showed that there are 17 million tons of forest industries residues and other wastes that can be delivered for under $30 per dry ton, our economic criterion. These quantities are sufficient for 520 ethanol plants, each of 10 million litres annual capacity. These sources of feedstock are spread across the country from the Maritimes to the West Coast. Thus it is scarcely conceivable that the development of a biomass based ethanol industry can be limited by available feedstocks. Futhermore, Clarke reported that there are 25 million hectares of undeveloped land and abandoned farm land suitable for planting fast growing trees or, as Hayes suggested,

energy crops such as Jerusalem artichokes or kochia, so these plants need never run short of feedstocks.

The total output of these plants, 5,200 million litres per annum, is about 20% of recent annual motor fuel consumption in Canada. It would be of some interest and importance to consider the social consequences of undertaking to build these plants. The total expenditure of about $4.0 billion would place such an undertaking in the class of a megaproject. It would bring a much needed boost to the steel, automotive, agricultural equipment, chemical equipment and forest industries, to engineers and to research. The employment consequences would be great. The small tax concession required to make the pump price of the ethanol so produced competitive with gasoline at today's pump price would be more than offset by increased government revenues in income taxation and decreased government expenditures on unemployment and welfare. Canada would indeed become energy self-sufficient, and these benefits would be spread across the country, providing a degree of regional self-reliance yet to be experienced in Canada.

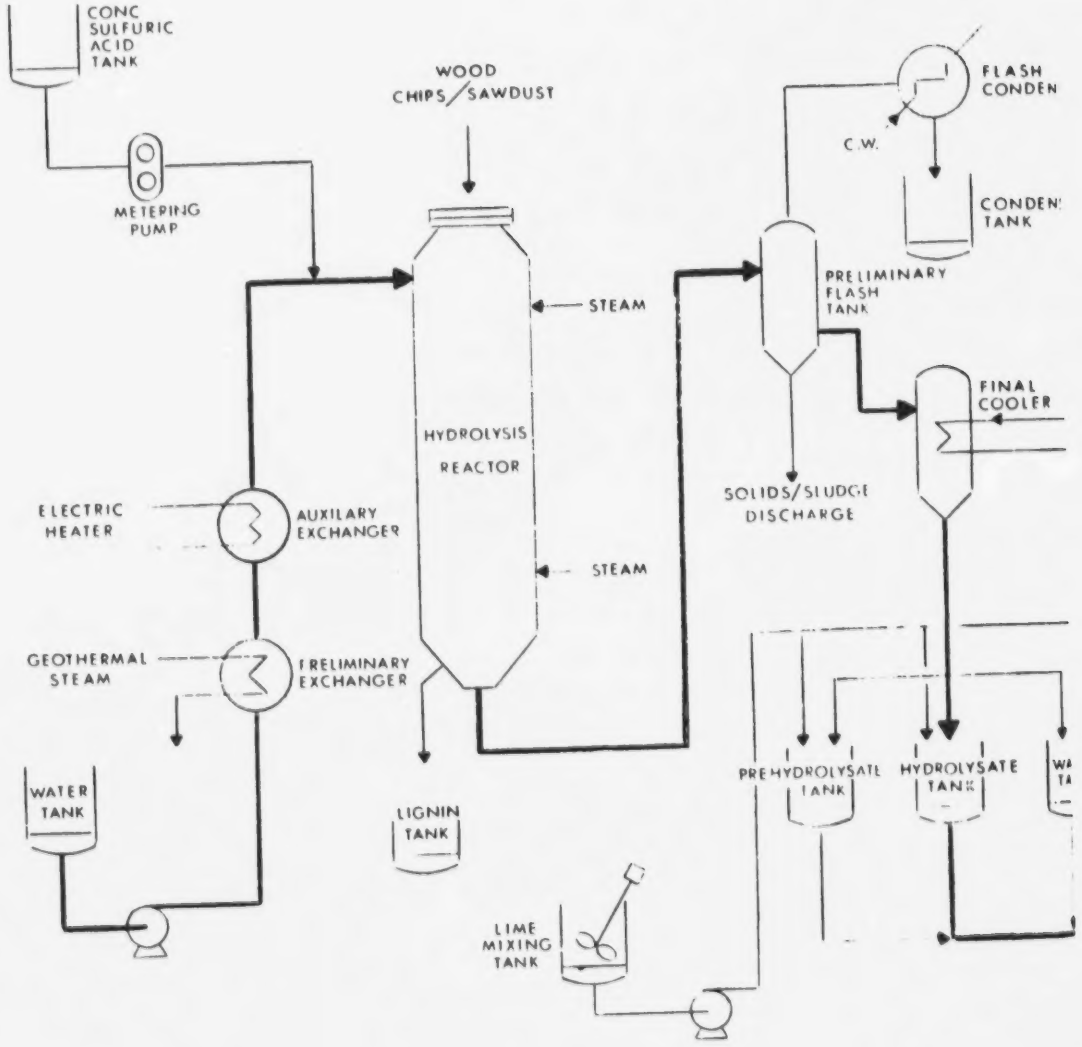
The payoff is great: let us hope some combination of the public and private sectors, an ambience in which Canadians are most comfortable, will pick up this golden opportunity and run with it.

When that happens, we can expect that each of the available technologies, including the FRI process, will have a contribution to make, and that over time one or a very few technologies will prove to be meritorious and become dominant.

6.0 APPENDICES

A. Process Flow Diagrams

B. Literature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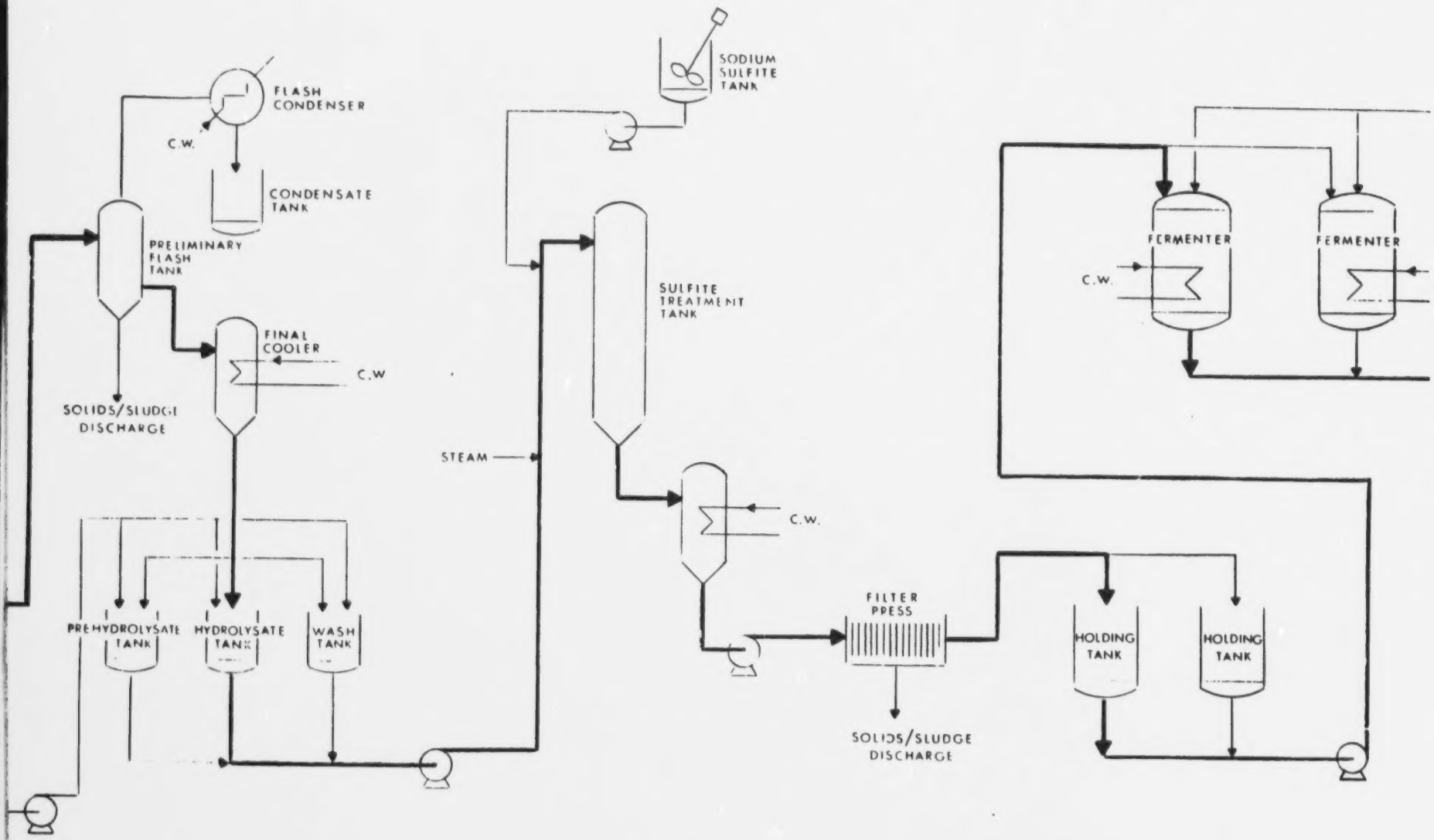
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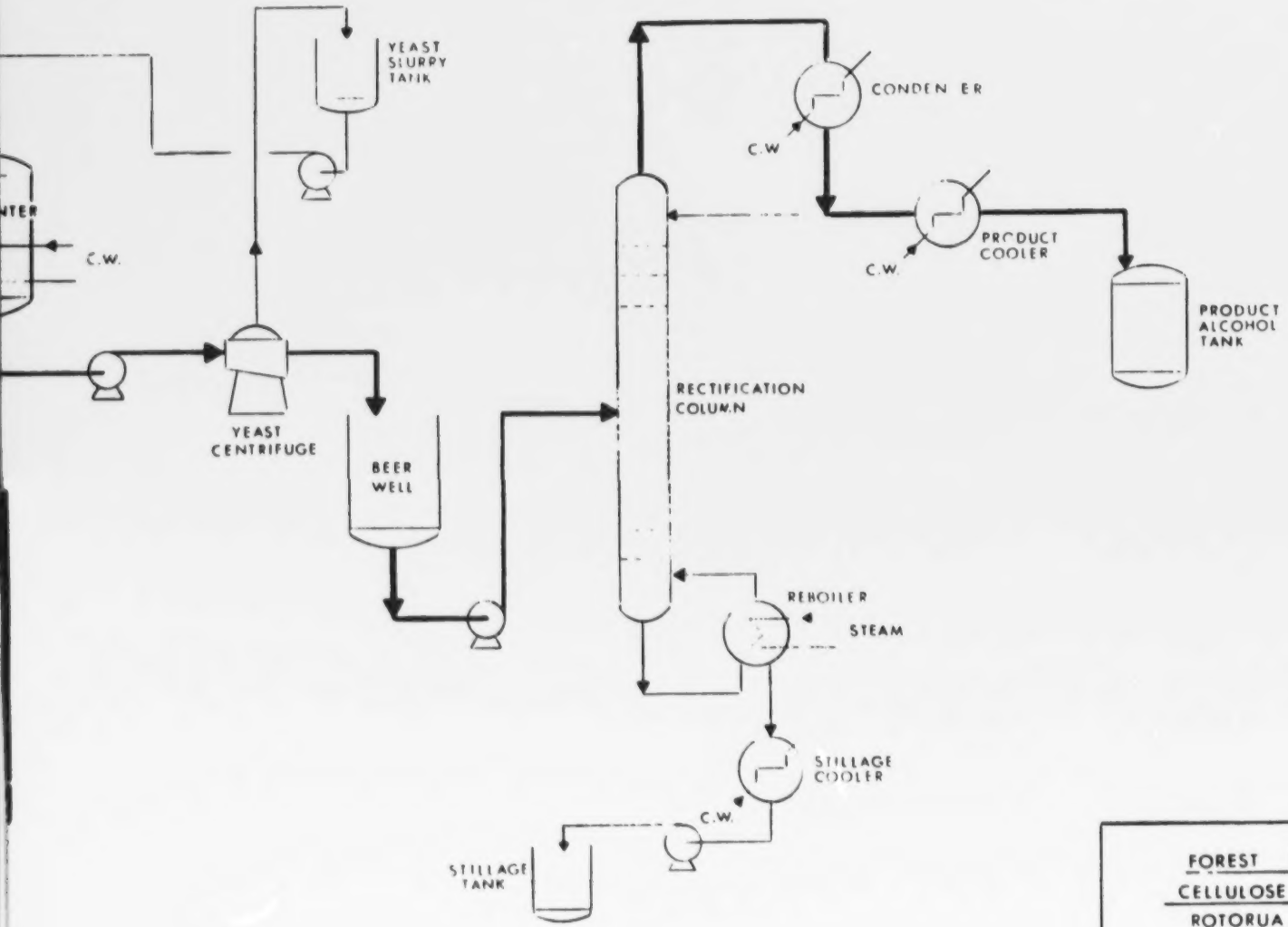

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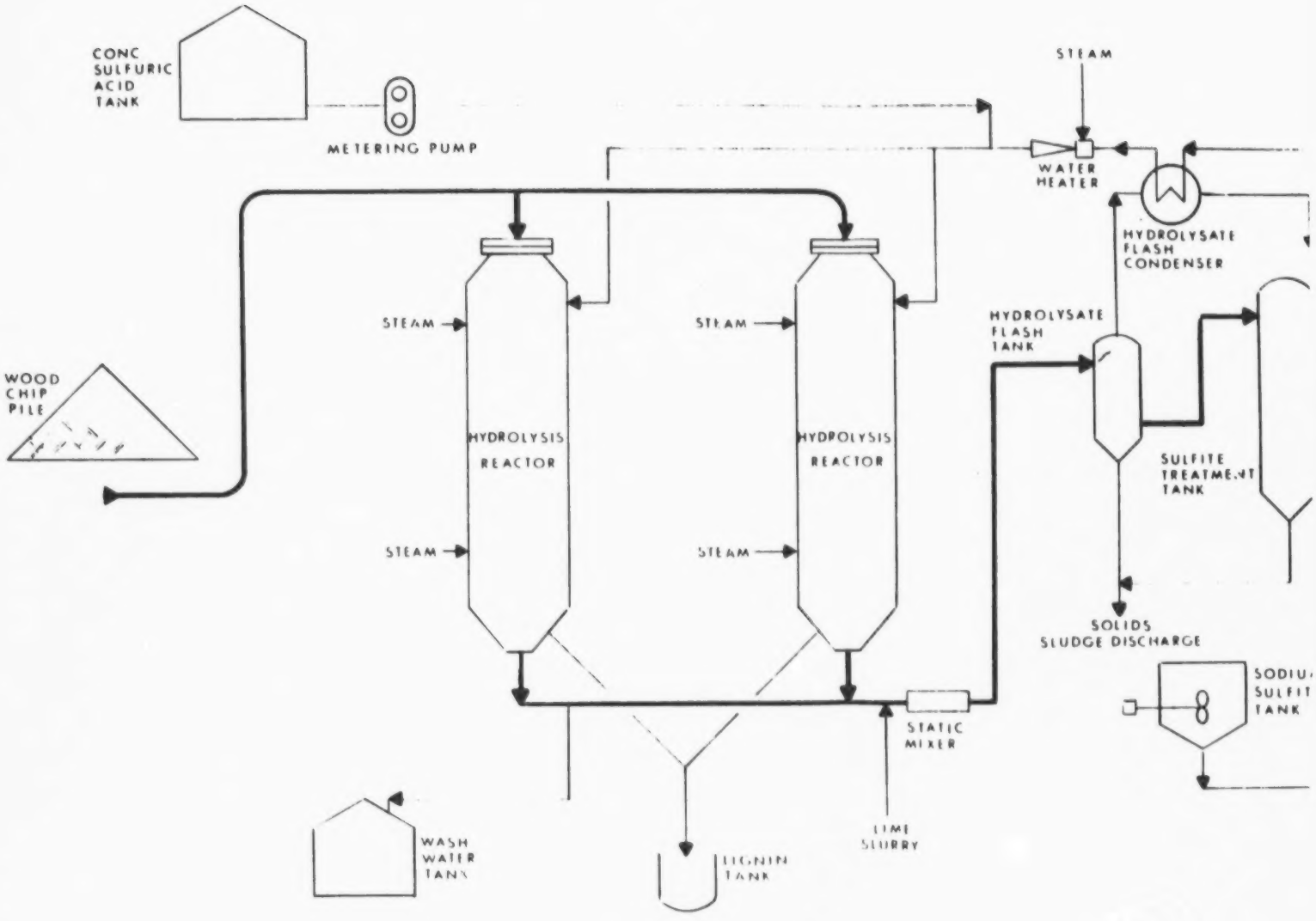
CELLULOSE TO ETHANOL PILOT PLANT

ROTORUA NEW ZEALAND

PROCESS FLOW DIAGRAM 1

OCT. 1982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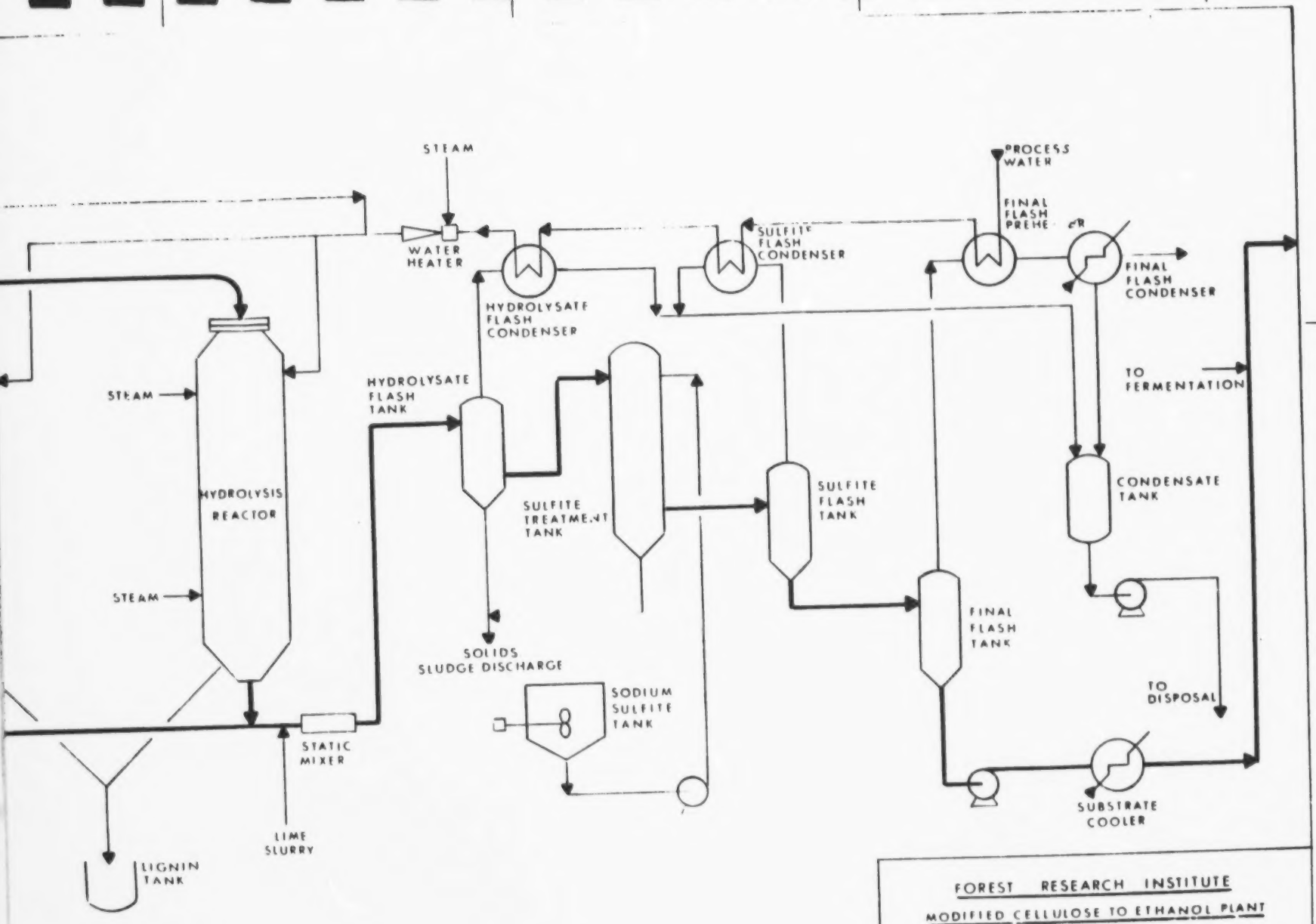

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MODIFIED CELLULOSE TO ETHANOL PLANT

30000 L PER DAY

PROCESS FLOW DIAGRAM 2

NOV. 1982

APPENDIX B

LITERATURE

PART I. Publications from the New Zealand Forest Research Institute

1. R. J. Burton, "The New Zealand Wood Hydrolysis Process", The Royal Society of Canada Symposium on Ethanol from Biomass, Winnipeg, October 13-15, 1982.

2. Keith Mackie, Kay Deverell and Ian Callander, "The Consequences of Wood Hydrolysis via the Dilute Sulphuric Acid Process", The Royal Society of Canada Symposium on Ethanol from Biomass, Winnipeg, October 13-15, 1982.

3. J. M. Uprichard and R. J. Burton, "Ethanol from Wood", Proceedings, Fifth International Alcohol Fuel Technology Symposium, Auckland, May 13-18, 1982, Vol. I, 317-324.

4. K. L. Mackie, "Wood Hydrolysis - Current Technology" Biotechnology Conference, Massey University, May 22-28, 1982, 15 pp.

5. D. A. Whitworth, "Wood as a Chemical Feedstock: Production of Ethanol", CHEMCA - 9th Australian Conference on Chemical Engineering, Christchurch, August 30 to September 4, 1981, 8 pp.

6. J. M. Uprichard, R. J. Burton, Kay F. Deverell, K. C. Mackie and D. A. Whitworth, "Ethanol from Radiata Pine via Dilute Sulphuric Acid Hydrolysis", OECD Workshop, Amersfoort, The Netherlands, October 8-10, 1980.

7. D. A. Whitworth, "Ethanol from Wood: Pilot Plant Operation", The Chemical Engineering Group, New Zealand Institution of Engineers, Dunedin, February, 1980, 23 pp.

8. "Liquid Fuel from the Forest", What's New in Forest Research, Forest Research Institute, No. 69, 1979, 4 pp.

9. D. A. Whitworth, "Production of Liquid Transport Fuel from Cellulose Material (wood):

"I Economic Considerations of Acid Hydrolysis of Wood for Subsequent Conversion to Ethyl Alcohol", New Zealand Energy Journal, November 25, 1976, 173-177.

"II Energy Conversion Efficiencies of the Processes", New Zealand Energy Journal, February 25, 1977, 14-17.

"III Laboratory Preparation of Wood Sugars and Fermentation to Ethanol and Yeast" (D. A. Whitworth and V. D. Harwood), New Zealand Energy Journal, October 25, 1977, 166-169."

10. Note Also: Nancy Phillips, "Wood Ethanol - Commercialization Underway Despite Controversy over Hydrolysis Techniques", Canadian Renewable Energy News, September 1982, p 34, 35, which comments on Derek Whitworth's construction of a new plant at Rainbow Falls, New Zealand.

PART II: U.S. Publications

11. F. Bergius, "Conversion of Wood to Carbohydrates", Ind. Eng. Chem. 29 249-253 (1937).

12. Sugars from Wood: a series of papers in Ind. Eng. Chem. 37 4-54 (1945), with these authors: E. C. Sherrard and F. W. Kressman, "Review of Processes in the United States Prior to World War II".

W. L. Faith, "Development of the Scholler Process in the United States".

E. E. Harris, E. Beglinger, G. J. Hajny, and E. C. Sherrard, "Hydrolysis of Wood: Treatment with Sulphuric Acid in a Stationary Digester".

J. W. Danning and E. C. Lathrop, "Saccharification of Agricultural Residues: a Continuous Process".

W. H. Peterson, J. F. Snell and W. C. Frazier, "Fodder Yeast from Wood Sugar".

R. H. Plow, J. F. Saeman, H. D. Turner and E. C. Sherrard, "The Rotary Digester in Wood Saccharification".

Jerome F. Saeman, "Kinetics of Wood Saccharification: Hydrolysis of Cellulose and Decomposition of Sugars in Dilute Acid at High Temperature".

R. L. Hasche, "By-Products of Wood Saccharification".

13. E. E. Harris and E. Beglinger, "Madison Wood Sugar Process", Ind. Eng. Chem. 38 890-895 (1946).

14. E. E. Harris, G. J. Hajny, M. Hannan, and S. C. Rogers, "Fermentation of Douglas Fir Hydrolyzate by S. Cerevisiae", Ind. Eng. Chem. 38 896-904, (1946).

15. E. E. Harris, "Wood Saccharification", Advances in Carbohydrate Chemistry, Vol. 4 p 153-188 (1948).

16. N. Gilbert, I.A. Hobbs, and J. D. Levine (TVA), "Hydrolysis of Wood Using Dilute Sulphuric Acid", Ind. Chem. Eng. 44, 1712-1720 (1952).

17. J. F. Saeman, "Production of Sugar and Alcohol from Wood", Report of a United Nations Seminar, Lucknow, October, 23 - November 6, 1952, 414-441. (In the same volume, p 193-198, K. N. Cederquist in "Some Remarks on Wood Hydrolyzation", describes his experiences in Sweden.)

18. D. F. Root, J. F. Saeman, J. F. Harris and W. K. Neill, "Kinetics of the Acid-Catalysed Conversion of Xylose to Furfural", Forest Prod. J.9 No. 5: 158-171 (1959).

19. S. W. McKibbins, J. F. Harris, J. F. Saeman and W. K. Neill, "Chemical Conversion of Wood Residues", Part V: Kinetics of the Acid Catalysed Conversion of Glucose to 5-Hydroxymethyl-2-Furaldehyde and Levulinic Acid" Forest Prod. J. 12 No. 1:17-22 (1962).

20. H. F. J. Wenzl, "The Chemical Technology of Wood", Academic Press, N.Y., 1970 Chapter IV, "The Acid Hydrolysis of Wood", 157-252.

21. (a) H. E. Grethlein, J. R. Lay and A. O. Converse, "Acid Hydrolysis of Cellulosic Material in Concentrated Slurries", A.I. Ch. E. Annual Meeting, Chicago, December 1976, 13 pp.

(b) H. E. Grethlein, "Acid Hydrolysis of Cellulosic Biomass", Manuscript about 1979, 19 pp.

22. B. Rugg and W. Brenner, (New York University), "Continuous Acid Hydrolysis of Waste Cellulose for Ethanol Production", Manuscript 1981, 10 pp.

23. H. R. Mendelsohn and P. Wettstein, (Inventa), Chem. Eng., June 15, 1981, p 63-65.

PART III: The USSR

24. "Hydrolysis and Wood Chemistry Journal", available in English translation from Allerton Press, Inc. An example which illustrates some aspects of Russian technology is:

L. N. Kraev, A. L. Fedorov, N. V. Horotkov, G. L. Blagushina, E. P. Merkulova and V. N. Churkin, "Industrial Testing of a Coniferous Wood Hydrolysis Process Using Hydrolyzate in the Cook".

Gidrliznya i Lesokhimicheskaya Promyshlennost, No. 6 pp 24-26, 1981.

25. N. I. Nikitin, "The Chemistry of Wood", Chapter XXIV, "Hydrolysis of Wood", (by B. I. Tokarev), pp 548-569. English translation 1966.